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THE STUDY OF INTEGRATED COAL-GASIFIER
MOLTEN CARBONATE FUEL CELL SYSTEMS

Contract No. 956389

FINAL TECHNICAL REPORT

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ABSTRACT

A novel integration concept for a coal-fueled coal gasifier-molten carbonate fuel cell power plant was studied. Effort focused on determining the efficiency potential of the concept, design, and development requirements of the processes in order to achieve the efficiency.

The concept incorporates a methane producing catalytic gasifier of the type previously under development by Exxon Research and Development Corp., a reforming molten carbonate fuel cell power section of the type currently under development by United Technologies Corp., and a gasifier-fuel cell recycle loop. The concept utilizes the fuel cell waste heat - in the form of hydrogen and carbon monoxide - to generate additional fuel in the coal gasifier, thereby eliminating the use of both an O_2 plant and a steam bottoming cycle from the power plant.

The concept has the potential for achieving coal-pile-to-busbar efficiencies of 50-59%, depending on the process configuration and degree of process development requirements. This is significantly higher than any previously reported gasifier-molten carbonate fuel cell system. The concept also offers natural resource savings, particularly under makeup requirements due to elimination of the steam cycle. In addition to continued development of the reforming molten carbonate cell, development of both low temperature catalytic gasification and high temperature desulfurization is required to achieve the high-end efficiency potential of the concept.

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1.0 EXECUTIVE SUMMARY

Coal gasifiers integrated with molten carbonate fuel cells have been shown to have promise for electric utility central station power plants. They have the potential for generating power at higher efficiency and with lower environmental intrusion than other base load fossil fueled power plants at competitive generation costs. In this study United Technologies investigated a novel integration scheme that improves the efficiency and environmental intrusion characteristics of coal gasifier fuel cell powerplants. This is achieved by using the waste heat from the fuel cell to drive the gasifier reaction. In conventional systems this waste heat is used to drive a bottoming cycle. By using waste heat in the gasifier, the bottoming cycle and the oxygen plant normally required in these systems is eliminated. The recovered energy is eventually returned to the fuel cell to be converted to electricity at a fuel cell efficiency which is higher than that of the bottoming cycle.

The objectives of the study were: to define a novel system integration scheme to improve system characteristics; to determine the overall efficiency potential of the novel concept and to compare it to previous studies and systems; and to define the component operating requirements and technology development necessary to achieve the efficiency potential of the concept.

The range of efficiencies for this integration concept is shown in Figure 1-A; the high-end efficiency is 59% (coal pile-to-busbar). This is significantly higher than previously reported efficiencies, also shown in Figure 1-A, for coal gasifier fuel cell power plants. Table 1-A shows in more detail the range of options studied and how their efficiencies fall within the range shown in the figure.

The configuration studies showed that water management (for preventing carbon formation and deposition) is a primary factor in achieving high system efficiency. As a result, the concepts that used the least energy for steam generation to prevent carbon yielded the highest efficiencies. The studies also showed the composition of the recycle stream has an important effect on the catalytic coal gasifier's operation. The composition is largely dependent on the method of fuel

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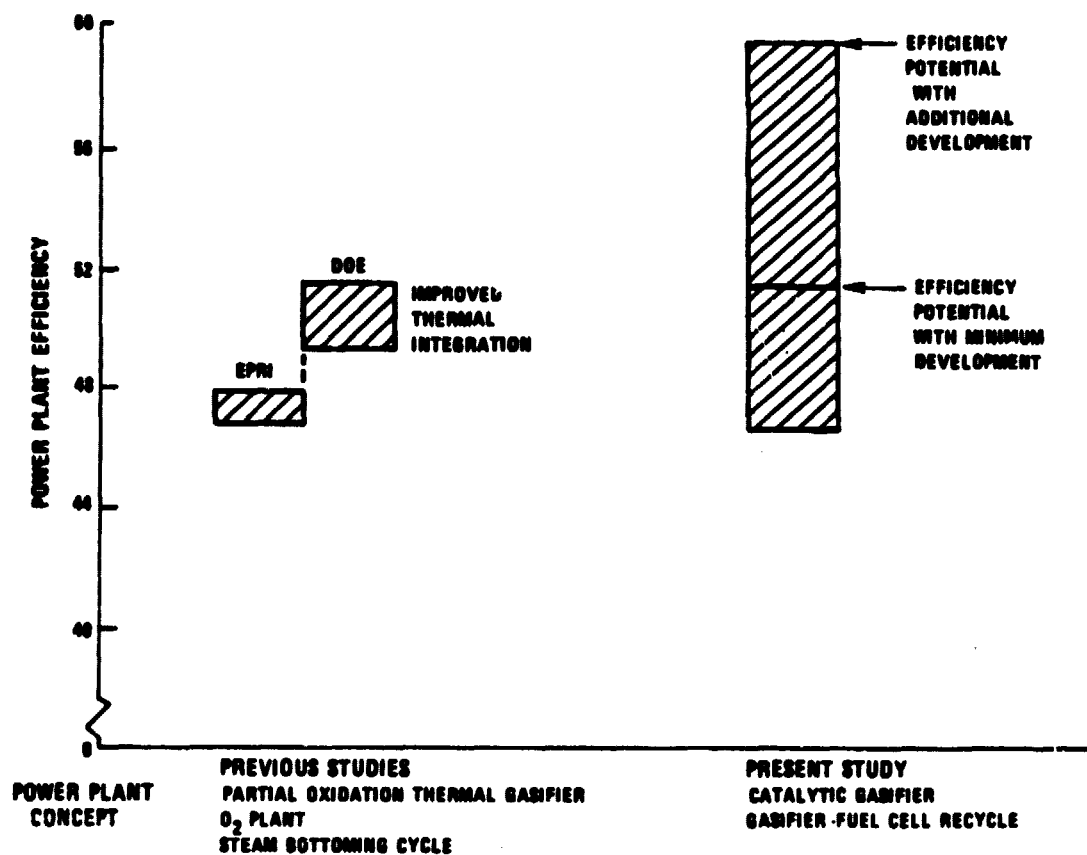


Figure 1-A. Efficiency Comparison for Range of Coal-Gasifier-Fuel Cell Power Plant Concepts

CONFIGURATION OPTION	ENHANCEMENT OPTIONS				
	1	2	3	4A	4C
SUBSYSTEM	BASIC				
DESULFURIZATION	LOW TEMP	HIGH TEMP	LOW TEMP	HIGH TEMP	
CO ₂ MANAGEMENT	ANODE VENT	PROD. EXCH. LOW TEMP	PROC EXCH. HIGH TEMP	SAME AS BASIC	SAME AS OPTION 3
FUEL GAS SATURATION	BOILERS	NOVEL	NONE		
WASTE HEAT REFORMING	INT. REFORMING				
CHARACTERISTICS				SLUR	SHR
EFFICIENCY (HHV COAL-PILE-TO-BUSBAR)	TO 40%	TO 60%	TO 52%	TO 47%	TO 52%
RELATIVE THERMAL REQUIREMENTS FOR RECYCLE SATURATION	HIGH	LOW	MODERATE TO HIGH	NONE	SAME AS OPTION 3
GASIFIER OPERATION	LOW TEMP		SIMILAR TO EXXON DEVELOPMENT EXPERIENCE	SAME AS BASIC	
COMPONENT DEVELOPMENT REQUIREMENTS (EXCEPT FUEL CELL AND GASIFIER)	NONE	DESULFUR- IZATION PROCESS	NONE	DESULFUR- IZATION CO ₂ PRODUCT EXCHANGE	HIGH TEMP RECYCLE BLOWER
HIGH TEMP AIR TO FUEL HEAT EXCHANGERS	YES	YES	NO	NO	

Table 1-A. Summary of System Configuration Characteristics

cell CO₂ management. For those configurations using a vent stream from the recycle loop, high CO₂ concentrations remain in the recycle gas. The CO₂ acts as a diluent and results in low gasifier operating temperatures. Removal of CO₂ via a separation process was an option that resulted in gasifier operation similar to Exxon development experience, as indicated in Table 1-A.

In general, the higher efficiency systems required more extensive process and component development than the less efficient systems. For example, development of high temperature desulfurization results in improved system efficiency for this integration concept. This is primarily due to its impact on water management. High temperature desulfurization eliminates loss of water in the recycle loop due to condensation, and the reduced saturation requirements yield higher efficiency potential.

With several configurations, system thermal management requires use of high temperature air-to-fuel heat exchangers. Although not considered to be development equipment, this type of exchanger is likely to be significantly more expensive than conventional devices. The determining factor is the level of leakage that is allowable in this application.

The efficiency difference between reforming fuel cell options, internal reforming, or sensible heat reforming was between 1 and 2 points. The loss was caused by added parasite power requirements for use in a second recycle to facilitate sensible heat reforming operation. In addition, high temperature blowers are required for the second recycle loop.

The results of this study indicate that further study is warranted. The configuration using commercially available desulfurization and CO₂ removal processes (Option 2) yields as high an efficiency as any previously reported. It provides good gasifier operating characteristics, moreover, and does not require high temperature air to fuel heat exchangers. It is recommended, therefore, that this configuration be further evaluated.

Development of methods for integrating hydrocarbon reforming with molten carbonate fuel cells, both the sensible heat and in-situ approaches, should be continued. The development results of both catalytic coal gasification and high temperature desulfurization should be monitored. Finally, the results of handling coal derived gases should be monitored in order to assure that the present understanding for prevention of carbon deposition is correct.

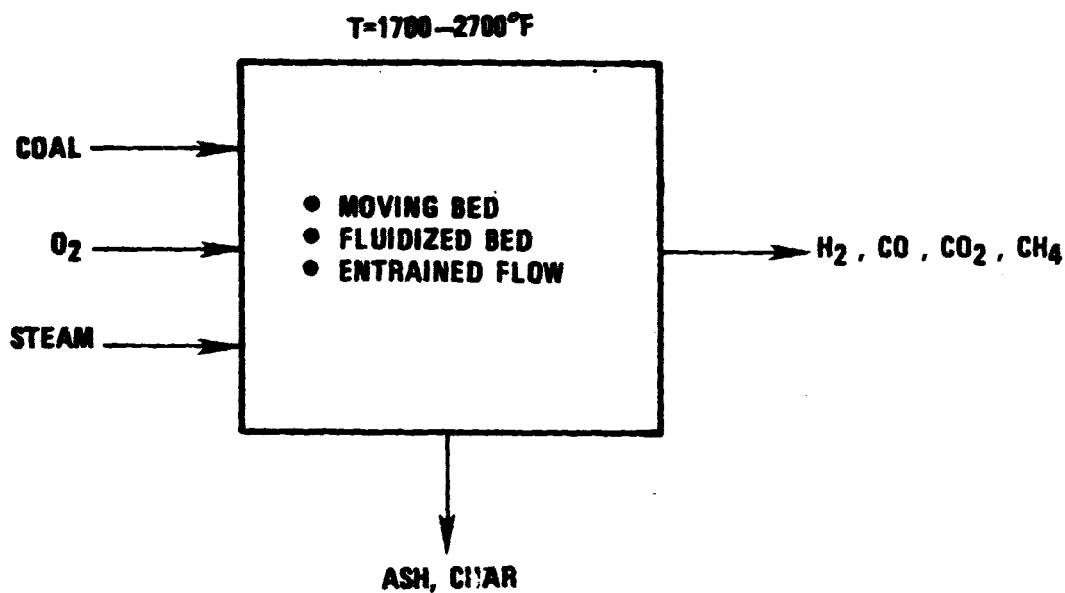
2.0 INTRODUCTION AND BACKGROUND

Studies have indicated that fuel cells can be integrated with a range of coal gasifiers into high efficiency central station power plants. This type of power plant concept is similar to that of an integrated coal gasifier combined cycle with the substitution of the highly efficient, direct electrochemical conversion fuel cells for the gas turbine Brayton cycle. Molten carbonate fuel cells are an especially attractive cell technology for this application because their operating temperature provides high grade waste heat for generation of steam that can be used for process or for power generation in a steam bottoming cycle.

2.1 DESCRIPTION OF CONVENTIONAL COAL GASIFICATION INTEGRATED WITH MOLTEN CARBONATE FUEL CELL SYSTEM

To date, the studies have focused primarily on integrating "conventional" type coal gasifiers with the fuel cell power section. Conventional coal gasification processes as represented functionally in Figure 2-A require a source of oxygen to supply process energy requirements. Since a portion of the coal or char is burned with oxygen or air, the heating value of the resulting gasifier product is lower than that of the original coal. This loss in heating value appears as sensible heat (increased temperature) in the gasifier product. The higher the gasifier exit temperature, the higher will be the hydrogen content of the gasifier product; the greater the amount of oxygen consumed and coal or char burned, the lower will be the gasifier efficiency. Generally, the lower the exit temperature of the gasifier, the higher will be the methane content and the higher the gasifier efficiency. A combined gasifier - fuel cell power plant system such as that shown in Figure 2-B will require an oxygen plant or oxygen supply system and a bottoming cycle to utilize the sensible heat energy created in the process, both from the gasifier as well as any high grade waste heat created by the fuel cell.

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DEVOLATILIZATION: $\text{COAL} + \text{HEAT} \longrightarrow \text{CHAR} + \text{HYDROCARBONS}$

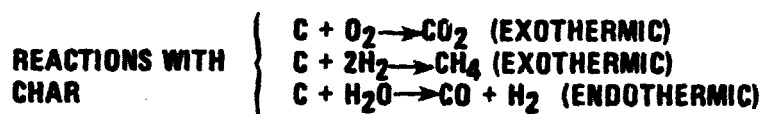


Figure 2-A. Conventional Coal Gasification

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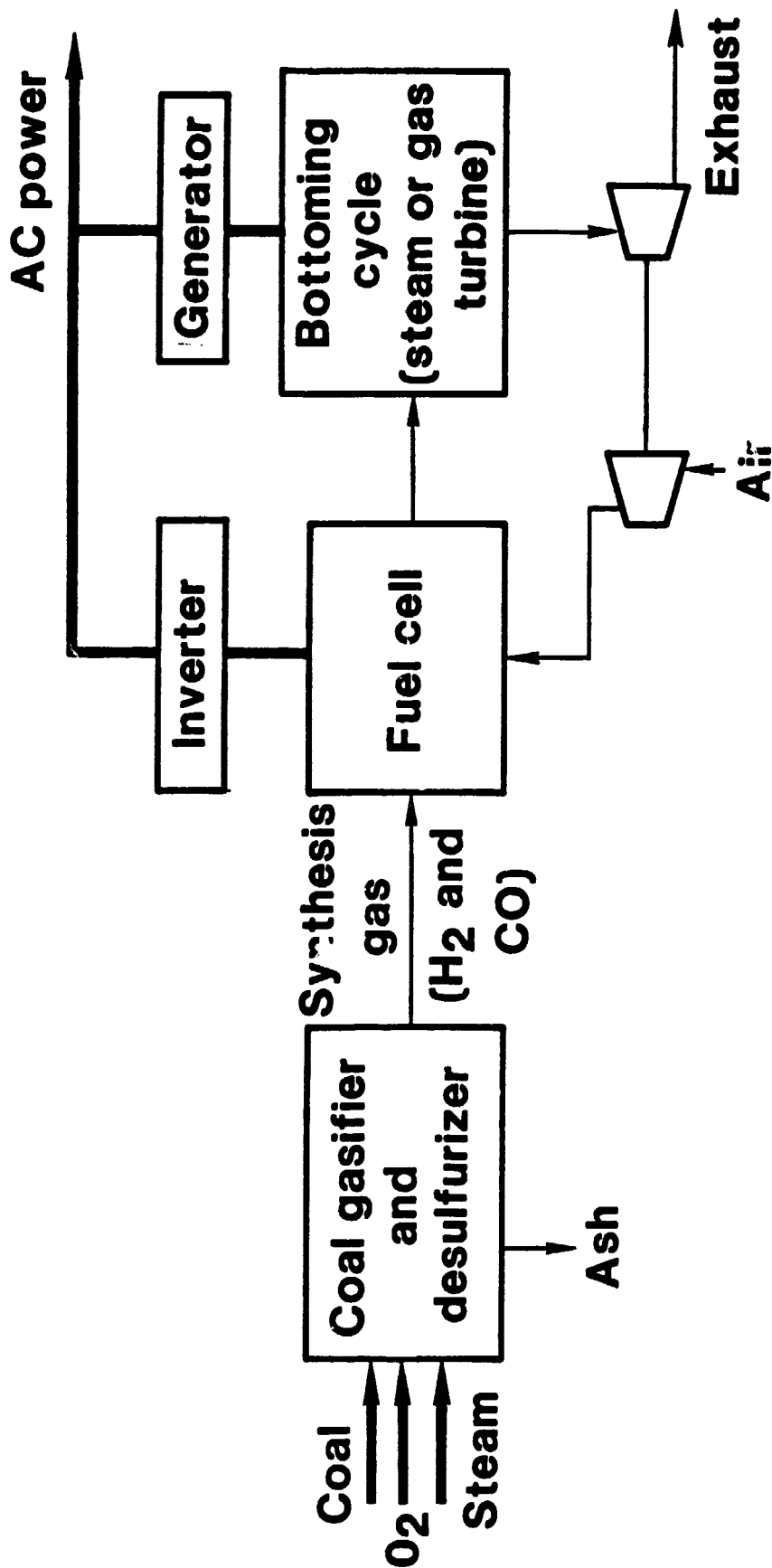


Figure 2-B. System Concept Central Station Coal Power Plant

The molten carbonate fuel cell offers a certain advantage over lower temperature fuel cells when it is integrated with a conventional gasifier whose product contains an appreciable methane content. In this case the fuel cell temperature is high enough so that a portion of the fuel cell waste heat can be utilized to steam reform this methane to hydrogen for direct consumption in the fuel cell. This "waste heat" reforming can be accomplished either through the process of "sensible heat" reforming external to the cell or "in-situ" reforming within the fuel cell package. These two waste heat reforming options are described in Section 4.1.4. This integration combines both the more efficient conventional gasifier with the fuel cell fuel conversion process in an improved coal gasifier-fuel cell power plant that reduces (but does not eliminate) system oxygen and bottoming cycle requirements.

2.2 DESCRIPTION OF CATALYTIC COAL GASIFICATION INTEGRATED WITH MOLTEN CARBONATE FUEL CELL SYSTEM

The use of a low temperature catalytic coal gasifier such as that described by Exxon^{1, 2} allows a degree of molten carbonate fuel cell system integration capable of producing high system efficiency without the need for a bottoming cycle or oxygen plant. Figure 2-C shows the functional requirements of a low temperature catalytic gasifier. Since no oxygen is required to sustain the reaction, the heating value of the product gas is maximized. (Since some char is produced in the process, its energy should be utilized to maximize process efficiency.)

In the catalytic gasifier the process pressure, composition, and temperature of the reactant feed gas affects the gasifier temperature and product composition. An increase in the feed hydrogen and carbon monoxide content forces the equilibrium toward more methane (exothermic), thereby raising the temperature. Additional steam shifts the equilibrium toward more hydrogen (endothermic), thereby lowering the temperature. Increasing gasifier pressure shifts the equilibrium toward more methane and higher temperature. Proper adjustment of the pressure and process feed conditions can result in a self sustaining gasifier operation at a temperature favorable to the kinetics of the catalytic gasifier process.

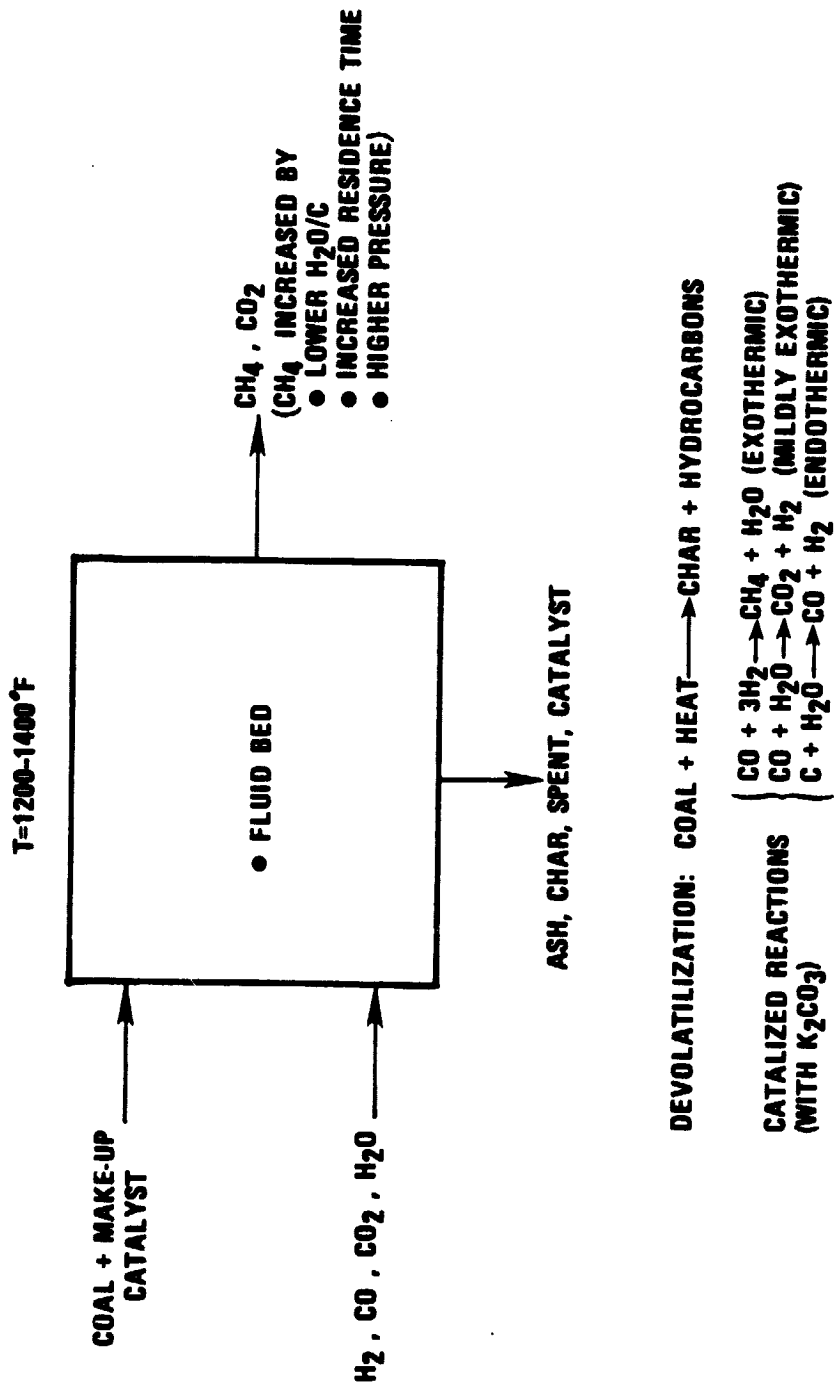


Figure 2-C. Catalytic Coal Gasification

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Combination of the molten carbonate fuel cell/waste heat reforming fuel conversion process with the low temperature catalytic gasifier results in a novel system which provides a means both of converting the gasifier methane product to hydrogen for the generation of electrical power and of supplying the feed and energy requirements of the gasifier. Most of the fuel cell waste heat is effectively utilized by "waste heat" reforming, thereby eliminating the need to recover this energy with a steam bottoming cycle.

Figure 2-D represents a simplified schematic of this novel system. Gasifier product is fed to the molten carbonate fuel cell/reformer. Fuel cell/reformer exhaust containing hydrogen is recycled to sustain the gasifier reaction. In this system it is possible to operate the gasifier at high pressure to favor the formation of methane and increase gasifier operating temperature and to operate the fuel cell at lower pressure to favor the production of hydrogen.

Figure 2-D represents only a simplified representation of this system. The detailed analysis of the important features of this system, such as water management, coal gas clean up, and product carbon dioxide removal, have resulted in a number of interesting system options that are discussed in this report.

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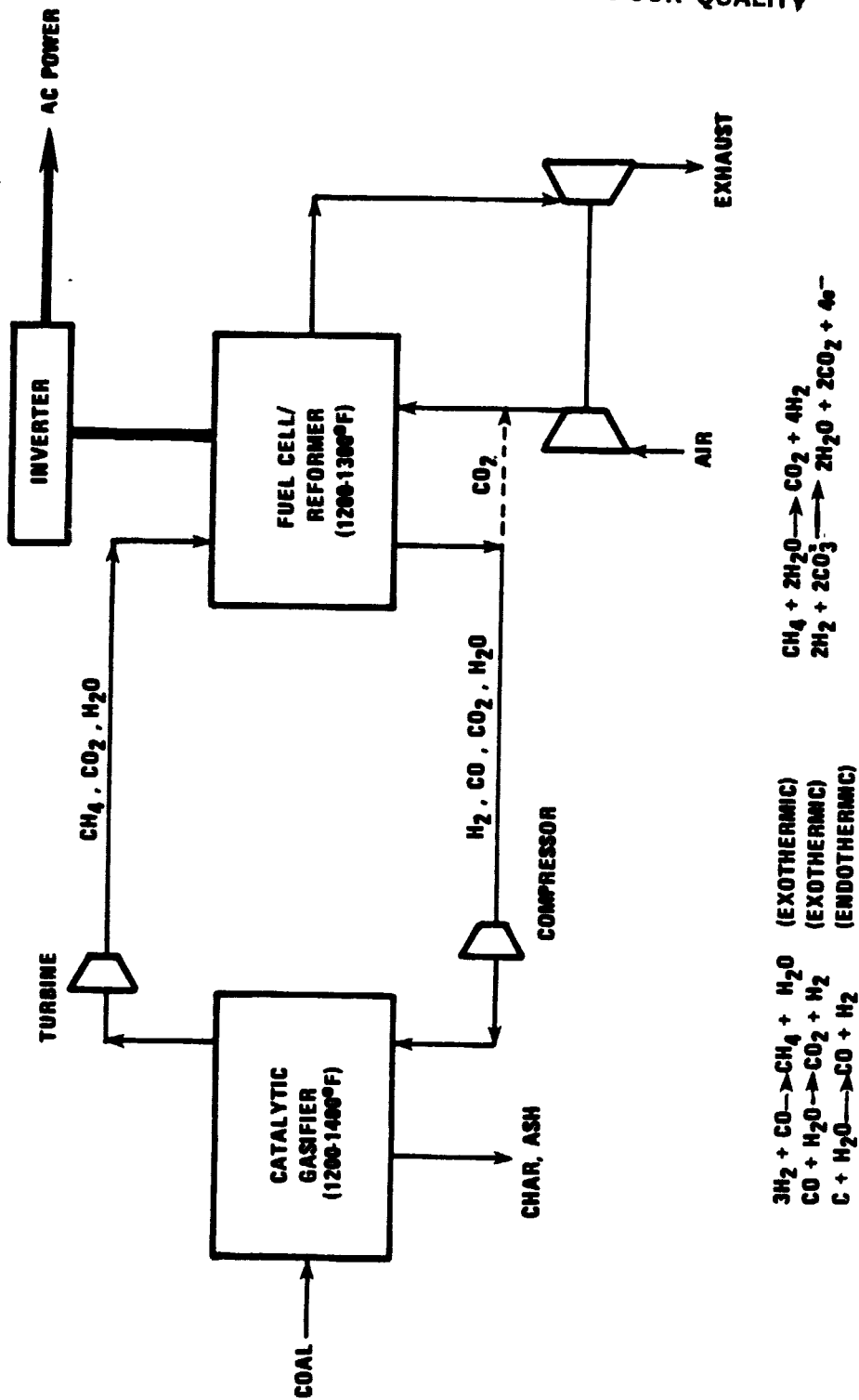


Figure 2-D. Integrated Catalytic Gasifier - Fuel Cell System Concept

3.0 STUDY, SCOPE AND APPROACH

The purpose of this study was to investigate the overall efficiency potential and the corresponding component design requirements of an integrated catalytic coal gasifier - molten carbonate fuel cell power plant concept. Major elements of the study include operation of the gasifier, the fuel cell section, and the gasifier/fuel cell recycle loop. No oxygen plant for the gasifier and no steam turbine bottoming cycle were used.

No economic estimates were generated in this study. Focus was on determining preferred process options for accomplishing the system integration, and modes of operation of the major processes and components, comparison of thermodynamic results, and recommendations for process development that would be required to achieve the system performance levels identified from this study.

The study approach consisted of two primary efforts:

- o Defining and evaluating a basic, or reference, system configuration.
- o Using the data from the basic configuration to identify and evaluate configuration options that might provide system enhancement.

The basic configuration is described in Section 4.2. It is based on integrating the Exxon-type catalytic gasifier with the fuel cell using commercially available process equipment. Study assumptions, especially including operation of the gasifier and fuel cell were established from publicly available literature and experience at Power Systems Division. These assumptions are reviewed in the following section.

Study evaluation parameters were established to focus on key component design requirements - especially the fuel cell and gasifier - as well as overall system efficiency. Sensitivity studies were then performed to determine the effect on the evaluation parameters for a wide range of study variables. The sensitivity studies were aided by use of computer programs that modeled the configuration.

Based on the results of this effort, several configuration options were identified to provide system enhancement. These enhancements took two forms: Options to reduce stringent component design requirements and options to improve system efficiency. Again, these configuration options were modeled by computer programs.

4.0 BASIS OF STUDY

4.1 GROUND RULES AND ASSUMPTIONS

4.1.1 Coal and Coal Feed

The coal type was specified by Jet Propulsion Laboratories to be Illinois #6, with an analysis as indicated in Table 4-A. Because this coal is identical to that used in several recent studies by United Technologies Corporation, it provides a good comparison between this integration concept and the other studies.

The coal preparation and feed process is based on design studies from Exxon's predevelopment program³. An understanding of the requirements of the process is important so as to be able to account for system thermal and electrical parasites and to determine the correct energy and mass balance around the gasifier. The steps assumed in the coal preparation and feed process include grinding, catalyst impregnation, coal drying, and the pressurized feed system. Figure 4-A shows these elements of the coal feed section in the overall system.

As-received coal is first ground to a particle size distribution optimal for the fluidized bed gasifier. The grinding is assumed to include an impact mill and classifying column to recycle the oversize particles with minimum fines formation. An inert gas blankets the equipment to prevent coal dust explosion, to elutriate the ground coal particles in the classifier, and to remove any rainwater absorbed during storage. The grinding stage of the coal preparation process is assumed to require ~3.3kW per ton of as-received coal.

In the second preparation step, coal is impregnated with potassium catalyst by continuous mixing of the ground coal with a catalyst-solution containing both recovered K_2CO_3 and makeup KOH. Potassium was found to be the most economical catalyst in the Exxon studies³. Two-thirds of the required catalyst is recovered from the char (See Section 4.1.3); one-third of the catalyst is makeup KOH. Based on Exxon's experience, the recycle K_2CO_3/H_2O catalyst solution is 30% by weight, and makeup KOH concentration is 20%. The assumed catalyst loading requirement is 15% by weight K_2CO_3 /dry coal, which results in a water/dry coal ratio of .55 by weight after catalyst impregnation.

TABLE 4-A

COAL ANALYSIS

<u>Type</u>	<u>Illinois No. 6</u>
PROXIMATE ANALYSIS (Wt %)	
Moisture	4.2
Ash	9.6
Fixed Carbon	52.0
Volatile Matter	<u>34.2</u>
	100.0
ULTIMATE ANALYSIS - DAF COAL (Wt %)	
Carbon	77.26
Hydrogen	5.92
Oxygen	11.14
Nitrogen	1.39
Sulfur	4.29
Other	<u>-</u>
	100.00
HEATING VALUE - AS RECEIVED	
High Heating Value (HHV) (Btu/lb)	12,235
Net Heating Value (LHV) (Btu/lb)	11,709

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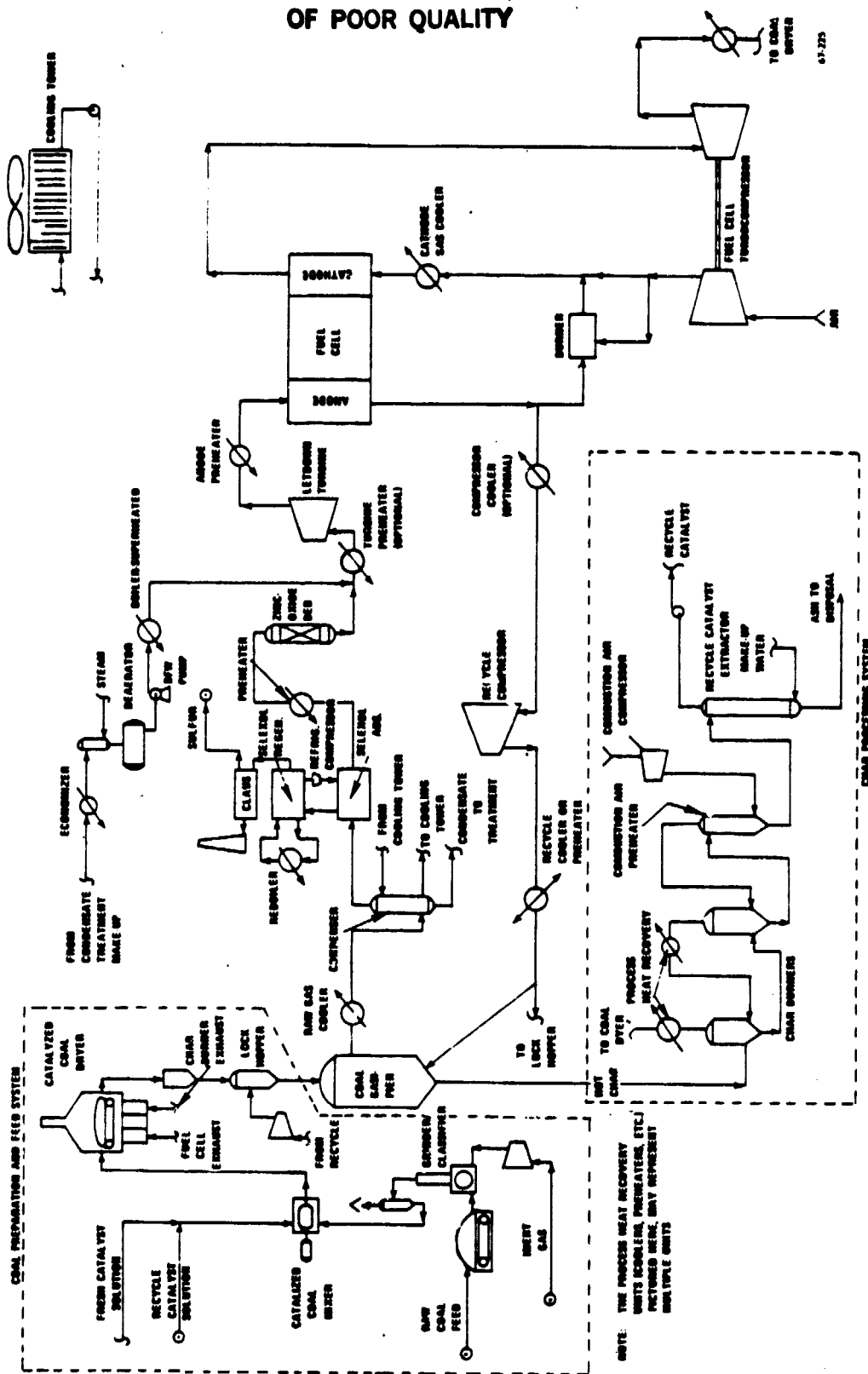


Figure 4-A. Basic Configuration Schematic

The wet catalyzed coal is heated to ~ 400°F and dried to 2.4 weight % moisture in the coal dryer by contact with warm flue gas. An ambient pressure dryer with low gas side pressure drop was assumed, although the details of the design were not a part of the study. Two flue gas sources are employed: the char burner flue gas and the fuel cell power plant exhaust gases. High quality heat is removed from these two sources, primarily for steam generation, prior to their use in the coal dryer. Coal drying is a major thermal sink in this gasification/fuel cell system concept, representing 5% of the heating value of the input coal. The use of low quality heat sources, however, mitigates the impact of this thermal requirement on system efficiency. Table 4-B summarizes the operating characteristics of the catalyst mixing and coal drying stages.

Drying of the coal with warm flue gases may provide a second desirable feature. The O₂-lean flue gases may prevent swelling of the coal by preoxidizing the coal prior to injection to the gasifier. Work done by Exxon⁴ indicates that coal swelling was responsible for lower than anticipated fluid bed densities, and that preoxidation with dilute O₂ gas streams brought the swelling phenomenon under control. In addition to O₂ concentration, residence time and preoxidation temperature also influenced the degree of swelling observed. Thus, drying with flue gas provides the potential for an O₂-lean gas stream whose inlet temperature and residence time with the coal could be adjusted in later design studies for optimum gasifier operation, with little impact on the remaining systems operation.

The last stage is the lock hopper system to feed the dry coal to the gasifier. Pressurization is accomplished with a bleed stream from the gasifier-fuel cell recycle loop.

4.1.2 Gasifier Operation and Modeling

The Exxon Catalytic Coal Gasifier is a fluidized bed gasifier that was under development by Exxon in the latter part of the 1970's and funded in part by the Department of Energy. The process uses alkali metal salts to catalyze the overall reaction of coal with steam to produce methane and CO₂ (See Section 2.2). The

TABLE 4-8

COAL PREPARATION

o RECYCLE CATALYST

 $K_2CO_3/H_2O = 30\%$ BY WEIGHT

o CATALYST REQUIREMENT

 $K_2CO_3/DRY\ COAL = 15\%$ BY WEIGHT

o COAL TO DRYER

 $H_2O/DRY\ COAL = 50\%$ BY WEIGHT

o DRYED COAL

 $H_2O/DRY\ COAL = 2.4\%$ BY WEIGHT

NET THERMAL DRYING REQUIREMENT

0.05 Btu/Btu COAL IN

catalyst serves to reduce agglomeration of caking coals. No tars or oils are produced. The process uses steam and recycled synthesis gas (H_2+CO), thus avoiding the conventional use of O_2 for partial oxidation of the coal. Major gasifier product components are CH_4 , CO_2 , unconverted steam, H_2 , and CO .

The dried coal is fed pneumatically into the bottom section of the gasifier. The gasifier is a single-vessel, single-bed reactor without special internals; it is anticipated that use of refractory linings will permit use of a low alloy steel shell. The steam and synthesis H_2 and CO recycle gases are injected through distributors to fluidize the solid char particles. This fluidization results in an essentially uniform bed temperature. A solids disengagement zone, located at the top of the vessel, and internal cyclones, minimize fines carryover. No external recycle of fines was assumed, although that should not impact the thermodynamics of the system. Ash, catalyst, and some unconverted char solids are removed from the bottom of the bed. In one Exxon scheme, the char-ash-catalyst solids are quenched with water and recycle synthesis gas prior to catalyst recovery; it was assumed for this study, however, that the hot char-ash-catalyst solids were pneumatically conveyed to the char burners for heat recovery (See Section 4.1.3).

The primary reactions occurring in the gasifier are the exothermic methanation and water-gas shift reactions together with the endothermic steam gasification reaction. Overall, the reactions are essentially thermoneutral when operated in the manner of Exxon development experience, with the preheated recycle and steam feeds serving as a heat source to offset the cooler-than-bed-temperature coal feed stream.

From the literature ³, the gases leaving the reactor bed have been observed to be close to shift and reform equilibria, with only traces of hydrocarbons heavier than methane. In this study, therefore, a simulation program to model gasifier operation was based on thermodynamic considerations only. Specifically, it is assumed that all the volatiles in the coal are gasified and that all the ash, catalyst, and some fraction of the carbon that is unconverted leave the bottom of the gasifier. Actual carbon conversions in the gasifier have been in the range of 90%; carbon conversion in this study was carried as a variable.

The gasifier simulation program iterates in order to solve simultaneously for mass and thermal energy balance around the gasifier based on achieving product gas shift and reform equilibria at a temperature approaching that of the gasifier bed. Thus, for a given coal feed, recycle feed and temperature, bed pressure, and carbon conversion, the program iterates in order to solve for the product gas equilibrium composition and temperature (corresponding to the bed temperature) so as to achieve overall heat and mass balance around the bed. To account, however, for the possibility that reform equilibrium is not attained, an approach temperature variable (TAPP) was introduced into the program, defined as:

$$TAPP = TEQ - TCG$$

where

TEQ is the temperature corresponding to reform equilibrium for the product gas stream,

and

TCG is the actual temperature of the bed and the product gas stream.

As reported above, the coal feed composition and temperature is assumed fixed in this study. Also, the system configuration and fuel cell operation determine the recycle feed to the gasifier; thus, gasifier variables include carbon conversion, bed pressure, recycle temperature, and approach temperature. The effect of the latter is to change the resulting product gas composition; for example, for a fixed set of gasifier variables, increasing the approach temperature increases the equilibrium temperature and, since the shift and methanation reactions are exothermic and inhibited by higher temperature, the steam gasification would be reduced.

4.1.3 Char Processing

The char processing section includes char burning, catalyst recovery, and ash disposal. The operational design is based on Exxon process development experience with some significant modifications.

Exxon's early predevelopment program focused on various aspects of catalyst use and recovery. Key findings from the study include:

- o Catalyst recoveries of $\sim 2/3$ could be obtained with water washing. This represents approximately 90% of the water-soluble potassium.
- o Recovery of approximately 90% of total potassium was demonstrated by use of a preferred chemistry sequence (calcium digestion).
- o Exposure of char to air inhibited catalyst recovery by oxidizing sulfides to sulfates.

From the above it was observed that heat recovery from the unconverted carbon in the char was considered, but the early results showed that catalyst recovery would be negatively impacted. In this study program, however, heat recovery by burning the char was assumed.

Figure 4-A also shows the char processing section. Several assumptions were made that impact the thermodynamics of this section:

- o Char contains only unconverted carbon, ash, catalyst, and sulfur; no coal volatiles.
- o Catalyst recovery is 67% with water washing only. The recycle solution is 30% by weight K_2CO_3 . (This defines the makeup KOH and the resulting drying requirements for the coal.)
- o The char contains sulfur at the same weight percentage as the parent coal (in this case, 3.9% by weight). The majority of the sulfur is volatilized primarily to H_2S in the gasifier. This precludes the necessity of SO_x treatment of the char burner flue gas.

In the first section of the char processing section, hot char from the gasifier is fed to a char burner where the unconverted carbon is combusted with preheated burner air. In this study the air feed was adjusted so that the flue gas temperature was approximately 1750°F, although the temperature could vary with design. The remaining solids - ash and catalyst and non-volatilyzed sulfur - is cooled via counter current flow from 1750°F to 200°F against fresh combustion air, which is heated from 200°F to 400°F in a simple moving-bed air preheater vessel. The hot flue gases from combustion of the char will be cooled to $\sim 425^\circ\text{F}$ in process heat recovery exchangers, with final cooling to 225°F in the coal drier.

Catalyst recovery is by water extraction. An agitated, multi-stage, countercurrent-flow extraction vessel is used, and full recovery of the soluble portion of the potassium salts is assumed. The effluent streams are the recovered catalyst solution (recycled to the coal feed section) and a "clean" ash slurry.

Disposal of the remaining ash should be relatively simple. It has been incinerated, cooled, and thoroughly water washed. The sulfur and potassium salts it contains are not water soluble. It is assumed that the ash will be drained of its (clean) water and landfilled.

The major parasite power consumers in this section are the combustion air compressor and the recycle catalyst pump, which are estimated to use ~28kW per ton of as-received coal, based on 90% carbon conversion. Lower conversions consume slightly more power, due to increased air flows. The net value of the recovered heat energy of the char at 90% carbon conversion is approximately 8% of the higher heating value of the as-received coal.

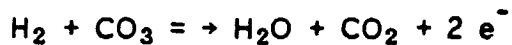
4.1.4 Fuel Cell Fuel Conversion Processes

The fuel cell fuel conversion processes include the molten carbonate fuel cell integrated with a reforming process. The reforming process serves to reform the methane-rich fuel gas from the gasifier to synthesis gas products (H_2 and CO) which are used by the fuel cell. The fuel cell electrochemically converts the H_2 to electrical DC power. Waste heat from the fuel cell reaction is used to provide the endothermic heat of reforming; thus the term waste heat reforming is used to describe this type of reforming concept. Several configurations for waste heat reforming are possible; two configurations were included in this study and are described in the following text.

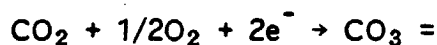
Fuel cells are electrochemical devices that convert the chemical energy of fuel gases directly into electrical energy. The application of the molten carbonate fuel cell was selected by UTC in this study for the following reasons: it offers high cell efficiency due to reduced activation polarization at high temperatures; it permits

waste heat reforming due to its high temperature; and it is actively under development at UTC.

The elemental molten carbonate cell is shown schematically in Figure 4-B. The cell consists of an anode, an ionically conducting electrolyte, and a cathode. Typically fuel gas, in the form of H_2 and CO and diluents such as CO_2 and H_2O are fed to the anode, where the electrochemical oxidation of the H_2 occurs as follows:



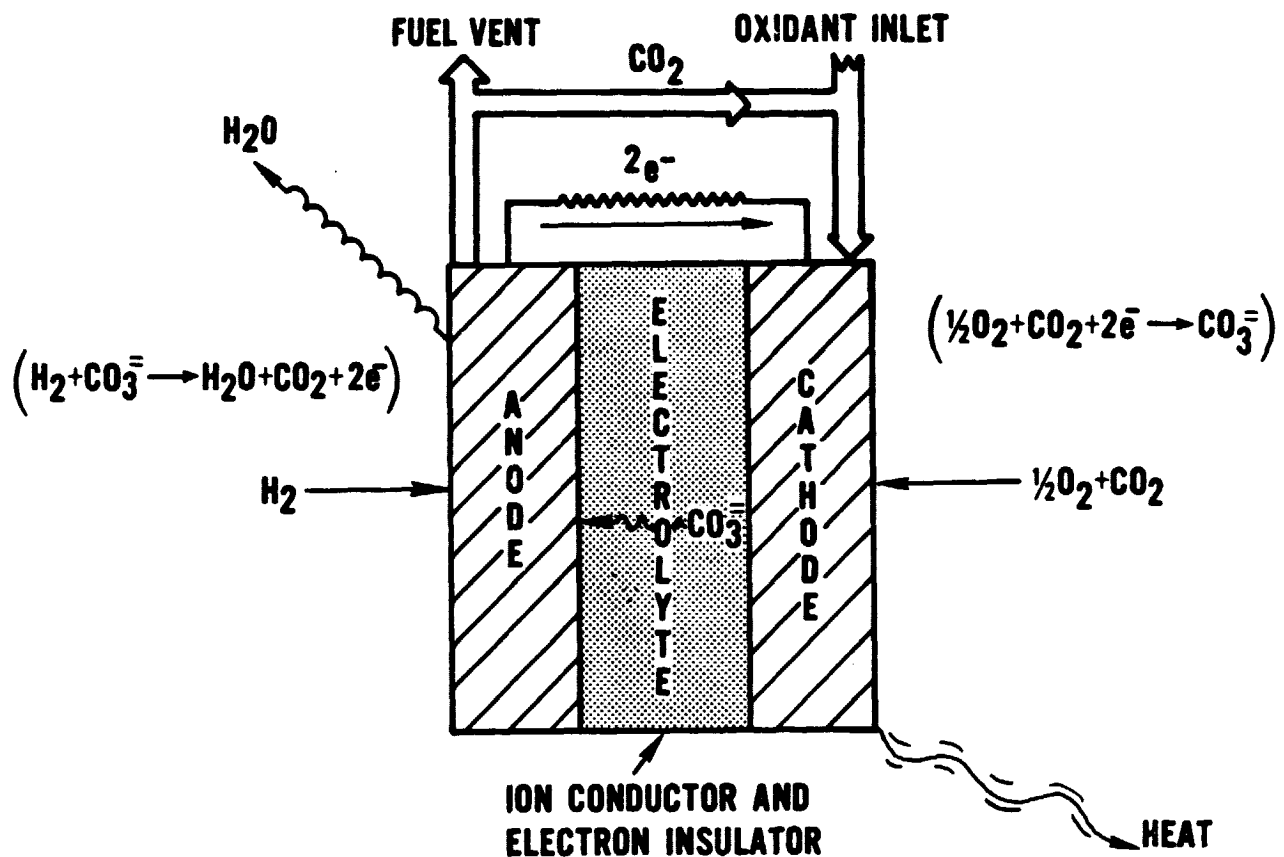
Simultaneously, CO is constantly being shifted in the cell anode compartment to make additional H_2 , so that the composition across the cell remains in water-gas shift equilibrium. Thus, the H_2 , either present in the inlet fuel gas or as a result of the water-gas shift, reacts with the carbonate ion CO_3^- to form byproduct H_2O and CO_2 , with an electronic current produced. The electrons are conducted through the load and back to the cathode. At the cathode, oxygen from air, and byproduct CO_2 from the anode reaction, combine electrochemically with the electrons to form the carbonate ion; viz:



The carbonate ion is conducted across the electrolyte and recombines with H_2 , completing the circuit. As indicated, the CO_2 formed at the anode must be transferred to the cathode to complete the cycle. This is accomplished via the CO_2 management approach, which either vents some or all of the anode exhaust gas to the cathode or extracts CO_2 from the anode exhaust stream. The CO_2 management approaches represent configuration options that were studied in this program. The overall cell reaction may now be written as:



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**ELECTROCHEMICALLY COMBINES H_2 AND O_2
TO RELEASE ELECTRICAL POWER DIRECTLY**

Figure 4-B. Elemental Molten Carbonate Cell

The byproduct of the overall reaction is H_2O which leaves as water vapor in the exhaust gases of the system, and heat.

The operating temperature range of the cell was nominally between 1100° and $1300^\circ F$. This permits waste heat to be used in the reforming process. Cell operating pressure was a variable in the study.

In addition to operating temperature and pressure, three other parameters define cell operation characteristics. These parameters include the fuel and oxidant utilizations and the cell performance. Reactant utilizations determine the variation in reactant partial pressure over the cell and therefore the ideal cell voltage and driving forces for reactant diffusion at each point in the cell. This is an important determinant of cell performance. Utilizations can be defined several ways. In this study, fuel utilization was defined as the ratio of the H_2 consumed in the cell to the theoretical H_2 available from the gasification process. Thus, fuel utilization is a partial measure of system efficiency since it indicates the percentage of the available fuel that is consumed electrochemically in the fuel cells.

The unit of performance for fuel cells is the design power density per square foot of active cell area. Cell power density is the product of the voltage measured across the electrodes of each cell and the current density of the cell at that voltage. Projected cell performance used in this study was based on an analytical model developed at PSD, which has been used in recent conceptual design studies^{5,6} for both EPRI and DOE.

Two approaches for waste heat reforming of the methane-rich fuel gas from the catalytic gasifier were included in this study. One of these is termed internal reforming, whereby the reforming process takes place in-situ in the anode of the cell. This option is depicted in Figure 4-C. This approach requires a catalyst activity at the anode sufficient to provide reforming of the gas as it passes through the anode. For this study, it was assumed that the fuel gas attains reform equilibrium as it passes throughout the anode.

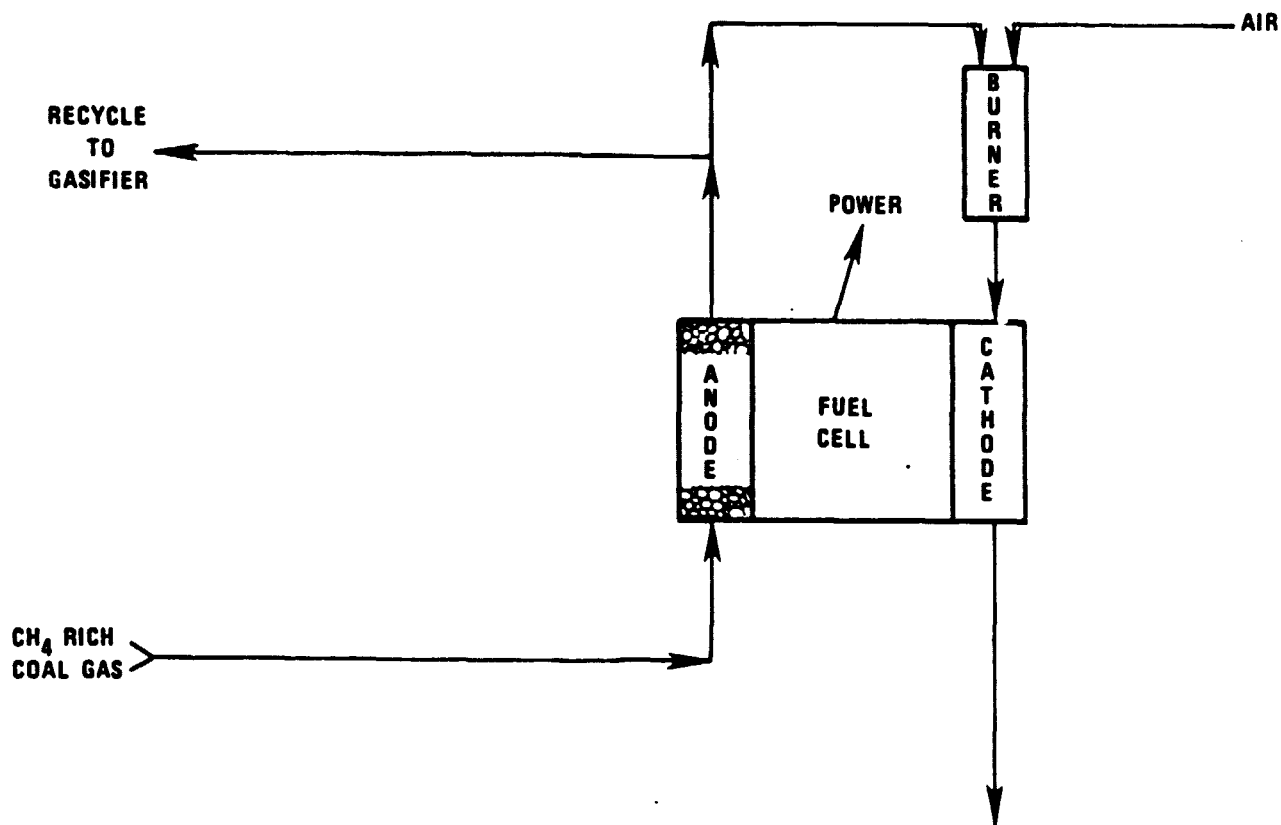


Figure 4-C. Internal Reforming Schematic

The second waste heat reforming approach is termed sensible heat reforming. With this option, the reforming process takes place in an adiabatic packed bed physically separated from the anode. Figure 4-D shows this option. The waste heat of the fuel cell electrochemical reaction is transported to the packed bed, typically by use of a recycle gas stream, where it provides sensible heat to the reform bed. Since the integration with the catalytic gasifier required a recycle anyway, it was thought that this recycle might be able to provide the sensible heat to the external reform bed for this approach.

4.1.5 Other System Assumptions

All the system options studied included operation of the gasifier and fuel cell fuel conservation systems as previously described. In addition, all the concepts include a fuel cell/gasifier recycle loop which carries the methane-rich gas to the fuel cell fuel conversion section and carries the unused synthesis gas back to the gasifier. This loop includes the process equipment to accomplish the following:

- o Fuel gas cleanup, specifically sulfur removal, down to levels of less than 0.1 ppm for delivery to the fuel cell
- o H₂O management to provide steam to the gasifier as well as to prevent carbon deposition in the recycle loop
- o CO₂ management to provide CO₂ to the fuel cell cathode

Approaches in the above three areas were varied, and, together with the waste heat reforming options, cover the range of the configuration options that were studied in the program. Each of these areas, therefore, will be covered in detail in the following discussions of the particular configurations.

The recycle loop also contains the necessary turbomachinery (recycle compressor and recycle let-down turbine, as required) to provide for recirculation and permit differential pressures for the gasifier and fuel cell. Operating temperatures and component efficiencies of these components were study variables. Major recycle pressure drops for all options were assumed to include 20 psia for the gasifier and 20 psia for the desulfurization section. Additional smaller losses were allowed for the heat exchanger and fuel cell fuel conversion components in the loop.

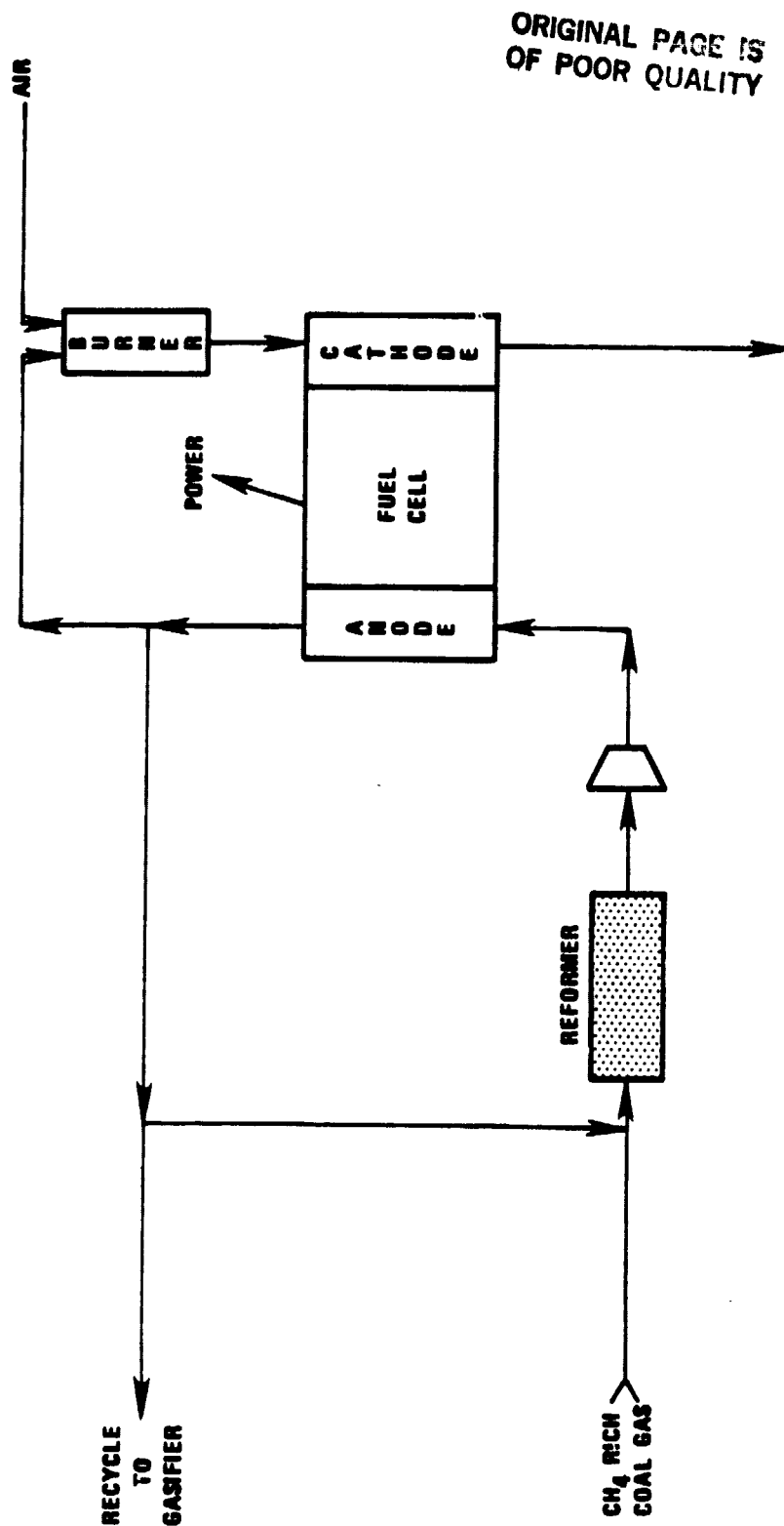


Figure 4-D. Sensible Heat Reforming Schematic

The system includes use of dry cooling towers for the removal of system waste heat, specifically the low quality heat of condensation in those configurations where H_2O must be condensed. The cooling tower fans and cooling water pumps amount to $3.25kW/10^6$ Btu of heat removal.

The fuel cell turbomachinery includes the air compressor for providing compressed air to the fuel cell stacks and burners as well as the expansion turbine for energy recovery. Again, the component efficiencies and the inlet temperature of the let down turbine were varied.

In some cases, excess shaft power was available from the fuel cell turbomachinery and/or the recycle turbomachinery. This was assumed to be converted into electrical power by use of AC generators operating at 98.5% efficiency. In other cases, electric motors with a 98% assumed efficiency were used to provide make-up shaft power to the turbomachinery.

4.2 BASIC SYSTEM DESCRIPTION

The starting point of the study was to develop a basic configuration scheme that integrated the Exxon-type catalytic gasifier with the fuel cell fuel conversion section. The premise for this basic configuration was integration in a process loop that used commercially available process equipment.

Figure 4-A is a schematic of the overall basic system configuration. Operation of the coal feed, gasifier, char processing, and fuel cell fuel conversion areas were covered in sections 4.1. The basic configuration assumes uses of an internal reforming fuel cell. Discussion here will focus on the process gas flow sections of the configuration, which include the process equipment in the gasifier/ fuel cell recycle loop and the fuel cell cathode process flow.

The raw product gas from the gasifier is cooled in a series of heat exchangers to the required H_2S removal temperature ($100^\circ F$). The gas can be cooled slightly below its dew point ($300-325^\circ F$) in a series of process heat recovery exchangers,

and to 100°F with cooling water. Virtually all water vapor in the product gas is thereby condensed and is sent to condensate polishing for boiler feedwater pretreatment.

The raw product gas enters the primary stage of sulfur removal which was assumed to be the low temperature, commercial Selexol[®] process. The H₂S and related sulfur compounds are selectively removed to ~ 2.0 ppm concentration in this stage. The product stream is reheated to ~ 700°F and flows through (commercial) zinc oxide beds where the sulfur concentration is further reduced to 0.1 ppm concentration. The Selexol[®] process operation is based on previous studies performed by UTC and others^{5,7,8} and assumes refrigerated operation. The major parasites are reboiler steam for regeneration and electric power for refrigeration compressors. Heat for the reboiler is provided by a closed cycle steam loop using process waste heat from the system. The amount of steam and electric power are primarily dependent on the pressure and the molar flow of the raw gas stream, and the level of sulfur assumed in the clean product gas. Elemental sulfur recovery is accomplished through use of a Claus plant, with tailgas cleanup. The zinc oxide bed is non-regenerable and must be periodically replaced.

The product gas exiting the zinc oxide beds is mixed with steam prior to expansion in a let down recycle turbine. Since the gas is essentially dry, steam re-saturation is used to prevent carbon deposition in the product gas steam prior to preheat to fuel cell inlet temperature. The criterion for steam requirements is fuel C-H-O equilibria at a boundary temperature of 800°F, assuming graphitic carbon. At temperatures below 800°F, it is assumed that kinetic limitations prevent carbon deposition, although it may be favored thermodynamically. These assumptions are consistent with results of studies performed by UTC in the RP-1085 program⁹.

Steam is provided by the water managements system, which is comprised of conventional condensate treatment, boiler feedwater pretreatment, and steam generation systems. In this system process condensate is collected, recycled to condensate polishing, and mixed with the required amount of comparably pretreated make-up boiler feedwater (BFW). The total BFW stream will flow through a series

of heat recovery steam generators that utilize process waste heat to preheat the BFW to saturation, boil, and superheat to 500°F for mixing with product gas. No allowance for boiler blowdown losses was included in the study.

The resaturated product gas passes through the letdown turbine and is preheated to approximately 1200°F using process heat recovery prior to the fuel cell. The pressure ratio of the turbine depends on the operating pressures of the coal gasifier and fuel cell, which were variables in the study. An optional turbine preheater also included in the study utilizes available high quality process waste heat to maximize turbine output.

In the internal reforming fuel cell, hydrogen is electrochemically consumed, and the shift and reform equilibria are maintained by converting CO and CH₄ to H₂. Product H₂O and CO₂ concentrations gradually rise across the cell. A portion of the cell net waste heat is removed as sensible heat in the 1300°F anode exhaust stream.

CO₂ management is accomplished by continuously venting part of the anode exhaust stream to the cathode. Unutilized H₂, CO, and CH₄ present in the vented steam are oxidized to CO₂ and H₂O in a burner. This approach represents the most commonly assumed CO₂ management approach in molten carbonate power plant studies.

The recycle gas is compressed before it returns to the gasifier. An optional cooler was studied to reduce compressor power requirements and to avoid compressor material concerns due to high temperature. When used, the gas is cooled to a temperature slightly above its dew point. The recycle compressor shaft power is provided by the recycle turbine. The high pressure recycle gas is preheated prior to entering the gasifier. This preheat temperature was a variable in the study; in general, however, this temperature was limited to the range of 1200°F by the thermal quality availability of process waste heat. It is possible that auxiliary fuel could be used to preheat this recycle stream to higher temperatures, but this was not considered in the scope of this study.

Air for the fuel cell is supplied by the fuel cell compressor. Part of this air flows through the burner to oxidize the vent stream. The air and burner exhaust streams are combined and cooled to the cathode inlet temperature of 1100°F. This cathode process stream provides CO₂ and O₂ for the cathode reaction, and serves to provide thermal management to the fuel cell by removing cell waste heat as sensible heat in the 1300°F exhaust stream. Some of the thermal energy is extracted from the hot exhaust gas by the fuel cell turbine. This turbine shares a common shaft drive system with the fuel cell compressor; the net surplus shaft power drives an AC generator.

Process heat is recovered from the turbine exhaust which is then mixed as needed with char burner flue gas for use in the catalyzed coal drier. The remainder is exhausted directly up the stack.

The fuel cell turbine and compressor are assumed to be single stage machines. While their thermodynamic efficiency could be improved by interstage reheating and cooling, respectively, the potential benefits would be offset by lower quality process heat availability and thereby increased difficulties in process heat integration.

4.3 EVALUATION PARAMETERS AND STUDY VARIABLES

Evaluation parameters were established for the study to permit assessment of the overall thermodynamic potential of the integration concept as well as the design requirements placed on the major components. These evaluation parameters include:

- o Thermal requirements for steam resaturation in the recycle for prevention of carbon deposition and for gasifier operation. Previous studies determined that the degree of resaturation could significantly affect thermal energy balances in the system and impact system efficiency.
- o Gasifier performance in terms of the resulting temperature of operation for thermal balance around the gasifier.
- o Fuel cell performance as determined by the cell power density, watts/ft² (WSF).

- o Heat exchanger operating requirements, specifically, the need to use high temperature air-to-fuel heat exchangers for process heat recovery. Although not evaluated in this study, the need for this type of process heat recovery may necessitate the use of leak-proof designs which, although commercially available, would be higher in cost than conventional gas-to-gas designs.
- o Overall system thermal efficiency, coal-pile-to-busbar, defined as the ratio of the net electrical output to the higher heating value of the fuel into the system.

The variables included in this study have been described in the preceding sections and are listed in Table 4-C for ease of reference. The variables are grouped according to the section of the power plant concept in which they are located. Also included are resulting values of these variables for the basic configuration reference design point. This reference design point resulted from sensitivity studies showing the impact of the study variables on the evaluation parameters and represents the high-end efficiency potential of the system. These sensitivity studies, the resulting component design requirements, and system efficiency potential for the basic configuration are covered in the next section.

TABLE 4-C
BASIC CONFIGURATION OPERATING VARIABLES

Operating Variables	Resulting Design Point Values
<u>Recycle Loop</u>	
o Recycle Ratio - Recycle Molar Flow/Vent Molar Flow	1.0
o Optional Use of Recycle Compressor Cooler	Yes - Dew Point Temp.
o Optional Use of Recycle Turbine Preheater	Yes - 900°F
<u>Gasifier</u>	
o Carbon Conversion in Gasifier	90%
o Gasifier Operating Pressure	500 PSIA
o Gasifier Approach-to-Equilibrium Temperature	25°F
o Gasifier Inlet Recycle Gas Temperature	1200°F
Operating Variables	Resulting Reference Conditions
<u>Fuel Cell</u>	
o Fuel Utilization (Overall) = $\frac{\text{H}_2 \text{ Consumed Electrochemically}}{\text{Theoretical H}_2 \text{ Available}}$	75%
o Cell Voltage	0.75
o Cell Operating Pressure	1200 PSIA
<u>Turbomachinery</u>	
o Component Efficiencies	80%

5.0 EVALUATION OF BASIC SYSTEM CONFIGURATION

5.1 THERMAL REQUIREMENTS FOR STEAM SATURATION

Steam resaturation of the recycle is required to prevent carbon deposition in the dry product gas after low temperature desulfurization and to act as a feed to the gasifier. Initially, steam requirements were defined on the basis of preventing carbon. It was determined that additional steam was not required for the gasifier and would further detract from system performance.

Thermal requirements for steam saturation (Btu/Btw coal feed) are primarily determined by the flow rate of the gasifier/fuel cell recycle loop, as shown in of Figure 5-A1. The recycle ratio is a measure of the flow rate of recycle, defined as:

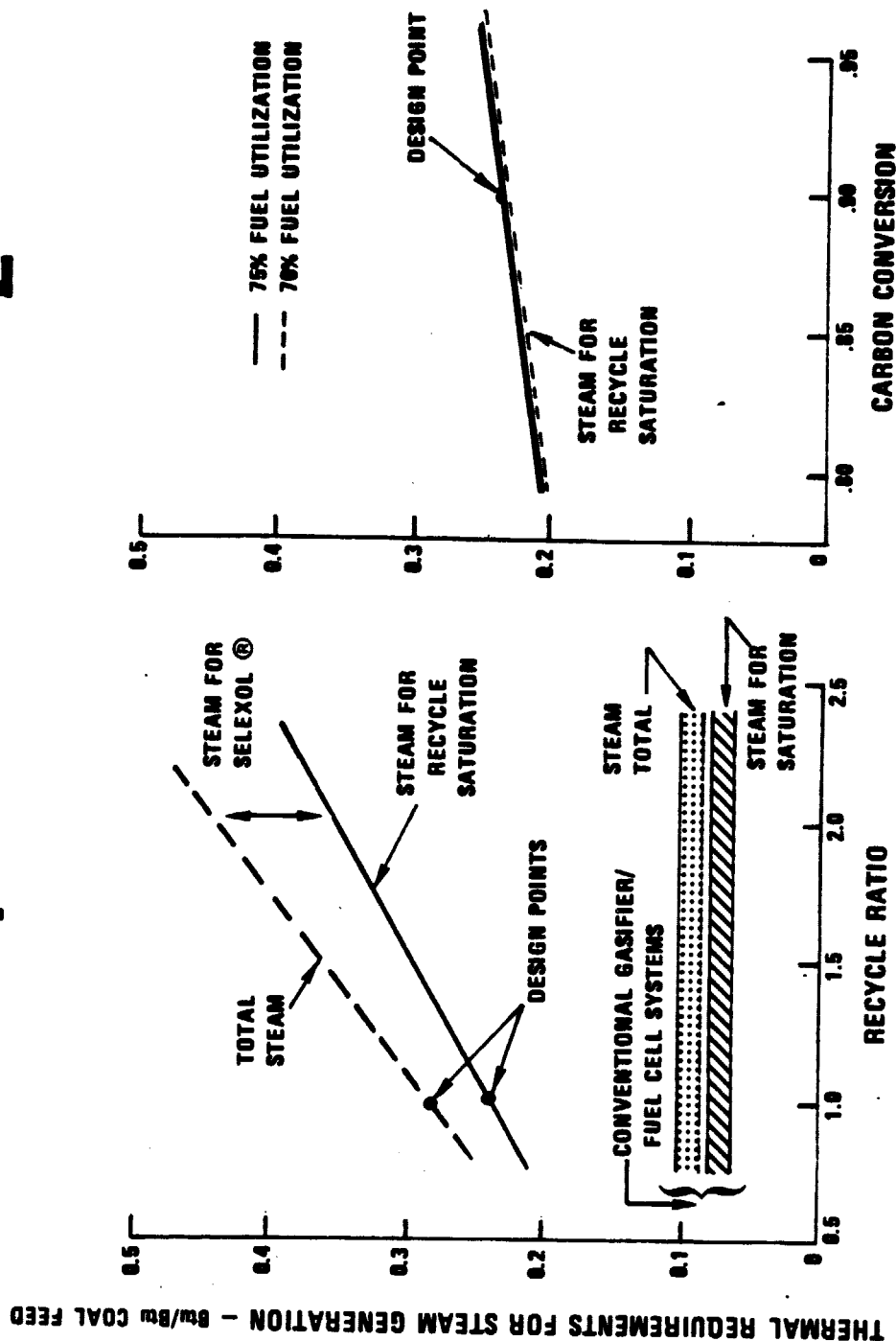
$$\text{Recycle Ratio} \equiv \frac{\text{Recycle Molar Flow Rate After Anode Vent}}{\text{Anode Vent Molar Flow Rate}}$$

As the recycle ratio increases, more H₂O is condensed upstream of desulfurization, and greater resaturation is required. As indicated in the figure over the range of recycle flow studied, the resaturation requirements to prevent carbon for the basic configuration are significantly greater than previous integrated coal gasification/molten carbonate fuel cell systems that had also employed low temperature gas cleanup. This increase is due to the use of the gasifier/fuel cell recycle and the resulting relatively higher flow of gases to the fuel cell.

Since the steam requirements to prevent carbon depend on full C-H-O equilibrium, it would be expected that in addition to recycle flow rate, only those study variables that change the relative C-H-O concentration of the recycle would change steam requirements. The two variables affecting this are carbon conversion and fuel utilization. The effects of these variables are shown in Figure 5-A2. Steam requirements increase with more carbon entering the system (higher carbon conversion), and decrease with higher fuel utilizations. The latter is due to the fact that the net mass products of cell reaction add O/C at a ratio of 3/1, which tends to inhibit carbon formation.

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Figure 5-A. Thermal Requirements for Basic Configuration

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5.2 GASIFIER PERFORMANCE

Comparison of the recycle flow rates and recycle composition to the gasifier between Exxon development experience and the basic configuration is shown in Figure 5-B. Two factors are indicated:

- o A recycle flow of 1.0 or less for the basic configuration is in the range of Exxon experience.
- o Although the water content based on carbon prevention is similar, the synthesis gas content ($H_2 + CO$) of the the basic configuration is significantly less than Exxon experience. The difference in composition is the CO_2 carried in the recycle stream as product of the fuel cell reaction.

The effect of the dilute synthesis gas feed to the gasifier is to reduce the exothermic methanation and shift reactions, thereby reducing available heat for the endothermic steam reforming of coal. Thus, for an assumed level of carbon conversion the operating temperature of the gasifier would be depressed relative to Exxon experience. Furthermore, for a fixed recycle composition, it would be expected that, increasing the gasifier temperature could be accomplished by bringing more sensible heat into the gasifier. This is shown in Figure 5-C, which plots the gasifier temperature as a function of recycle ratio (flow rate) and recycle temperature. Of significance is the low resulting temperature range for the basic configuration relative to the 1250 - 1300°F range for Exxon experience. Reducing fuel utilization, which increased the fractional amount of H_2 and CO in the recycle available for shift and methanation tends to increase gasifier temperature. A combination of high temperature recycle (1500 - 1600°F) together with high recycle ratios and low fuel utilizations is required to put the temperature of operation in the range of Exxon experience. It will be seen that high recycle ratios and low fuel utilizations significantly reduce system efficiency potential (See Section 5.4). Furthermore, there are no process heat sources in the system to raise the recycle temperature above ~ 1200°F. For example, char burning is a source of high quality heat but not of sufficient quantity for the recycle flow. A fired heater using supplemental fuel could be used but was not considered in the scope of this study. The required operating temperature of a catalytic gasifier would have to

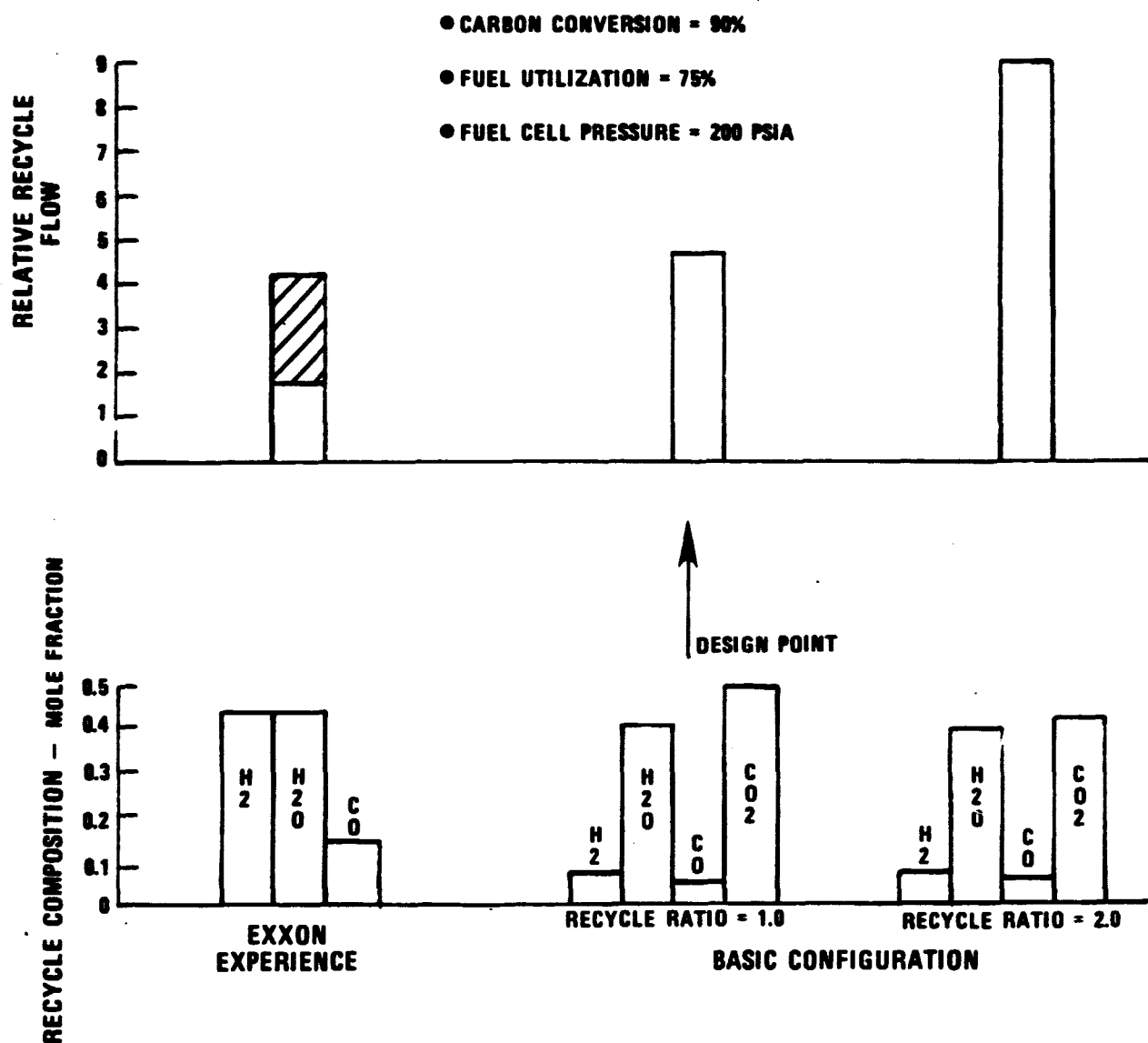


Figure 5-B. Gasifier Recycle Flow and Composition Comparison for Basic Configuration

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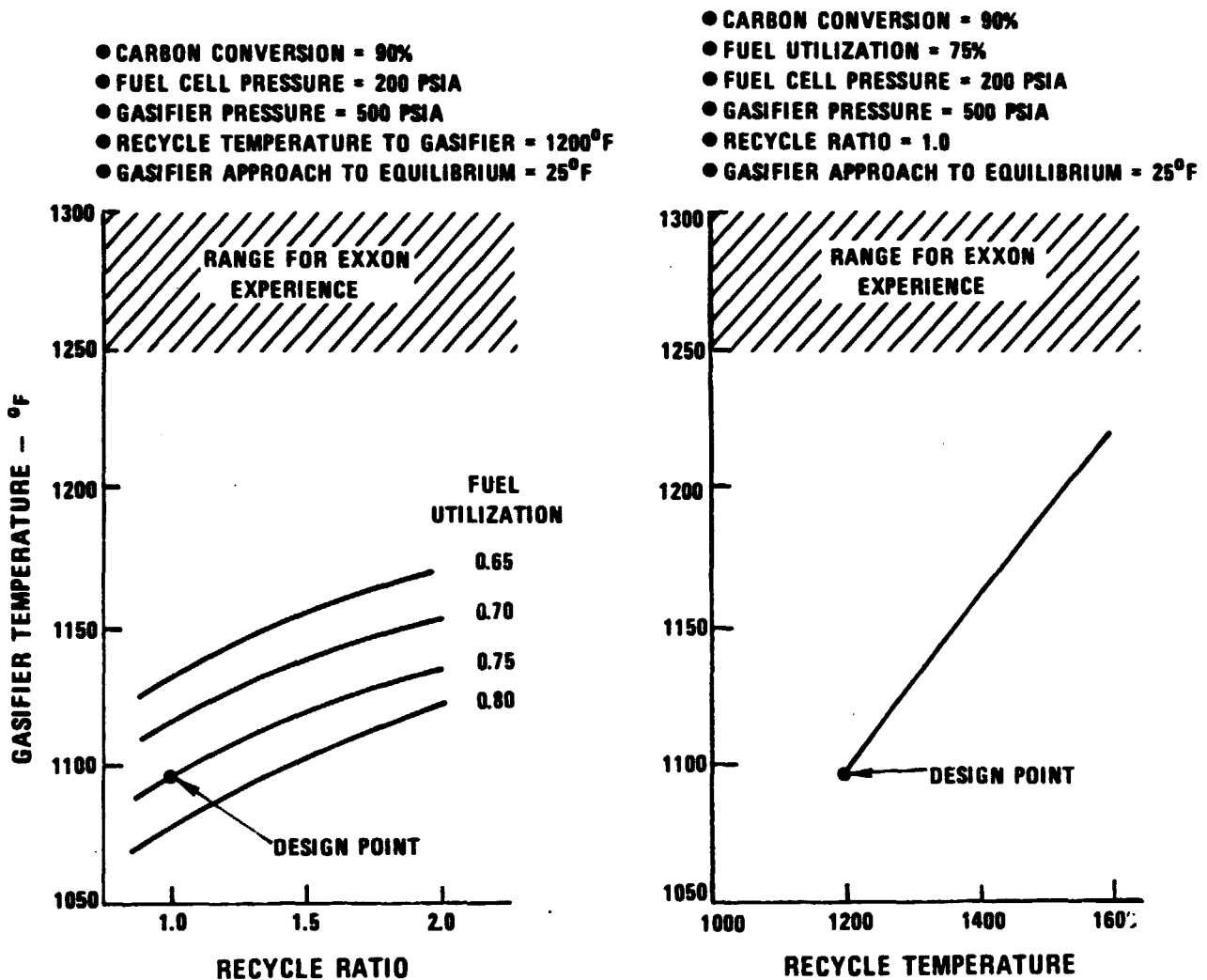


Figure 5-C. Gasifier Operation for Basic Configuration

be significantly reduced relative to present experience for the potential of the basic configuration to be realized.

5.3 FUEL CELL PERFORMANCE

Projected cell power density for the basic configuration is shown in Figure 5-D as a function of cell voltage and fuel utilization. Cell operating pressure for this map is 200 psia, although the latter has only a small effect on cell performance as shown in Figure 5-E.

The assumption of an internal reforming cell having a close approach to shift and reform equilibria across the cell anode yields comparable or slightly higher performance than previous recent studies. The latter were in the range of 145-160 WSF at cell voltages of .75 to .81 and utilizations of 80-90%. Thus, the conditions for fuel cell operation in the basic configuration do not result in dilute fuel gases and do not impose additional design considerations on the cell relative to present directions.

5.4 SYSTEM EFFICIENCY STUDIES

A major factor in overall power plant efficiency is the design operation of the turbomachinery, especially the fuel cell turbine operation. This can be seen in Figure 5-F1, which plots the net fuel cell turbocompressor output as a function of the fuel cell turbine inlet temperature. Fuel cell operating pressure is 200 psia. The design point for the basic configuration was selected to be 1300°F, which is the design exit temperature of the power plant exhaust from the fuel cell cathode.

Similarly, Figure 5-F2 shows the effects of the recycle turbocompressor design conditions on system operation. The significant variable is use of the compressor cooler, which assumes cooling of the anode recycle to a temperature just above the steam dew point. A compressor cooler also eliminates the concerns of high temperature compressor materials requirements. The use of a cooler was adopted as part of the reference conditions for the basic configuration. The use of recycle

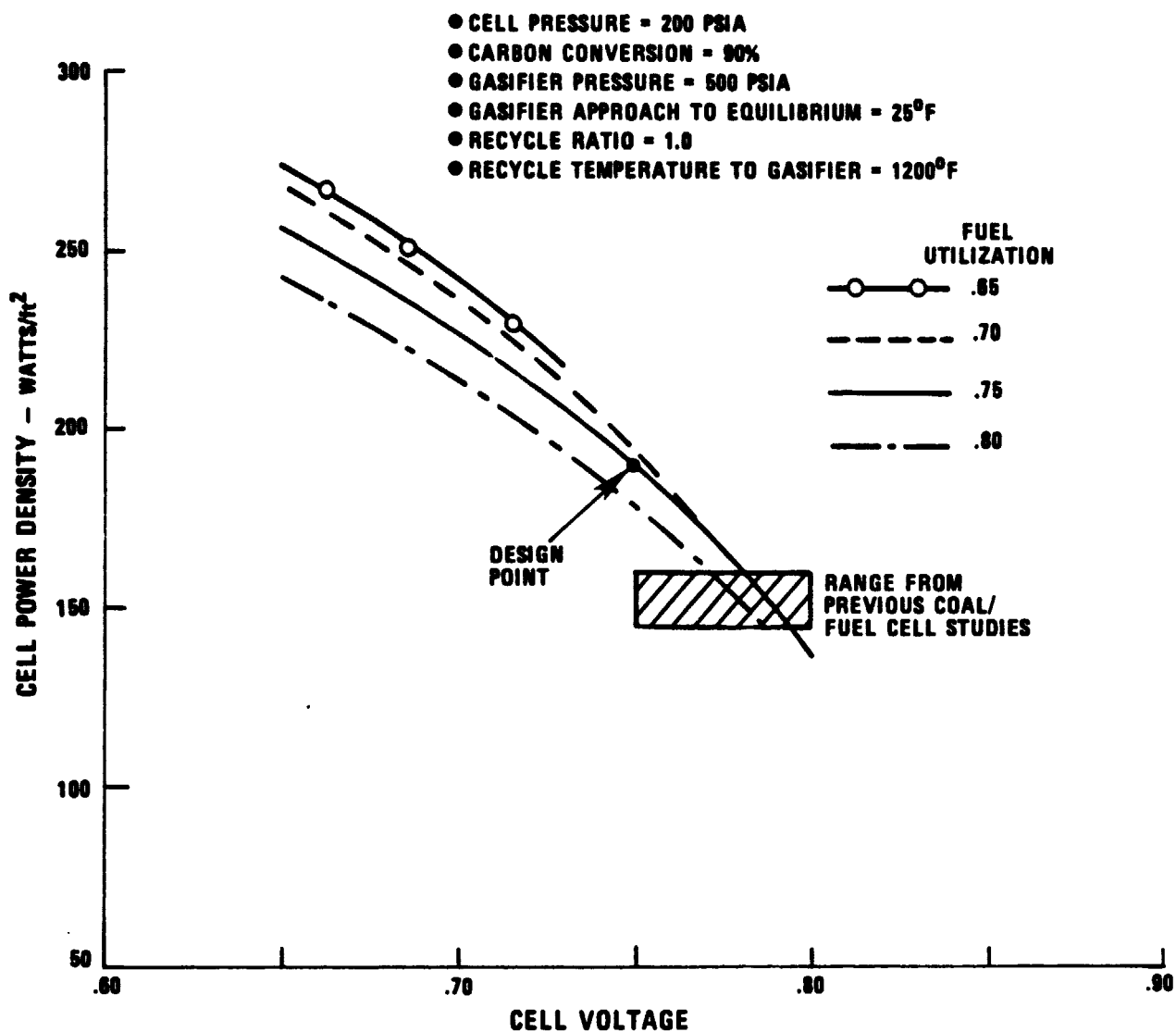


Figure 5-D. Basic Configuration Fuel Cell Performance Map

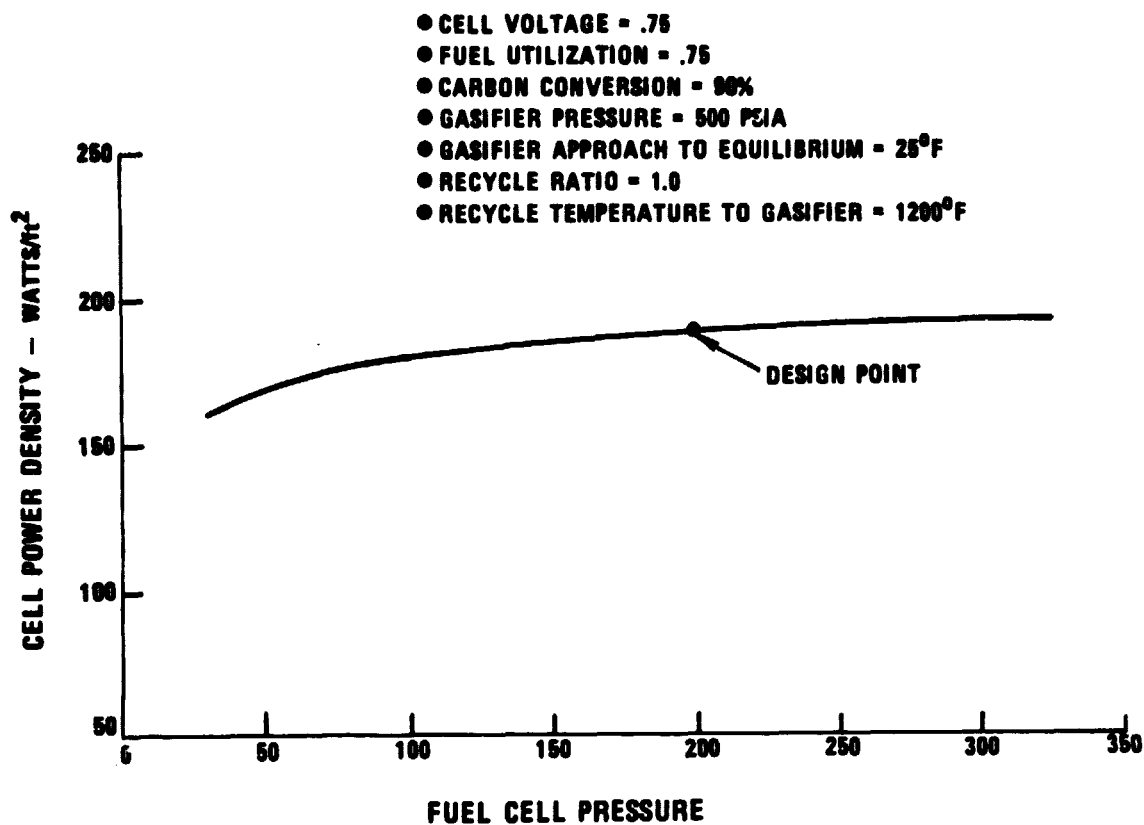


Figure 5-E. Fuel Cell Performance vs. Fuel Cell Pressure for Basic Configuration

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- FUEL CELL PRESSURE = 200 PSIA
- GASIFIER PRESSURE = 500 PSIA
- RECYCLE RATIO = 1.0
- CELL VOLTAGE = .75
- FUEL UTILIZATION = .75
- OXYGEN UTILIZATION \cong 67%
- TURBOMACHINERY COMPONENT EFFICIENCIES = 80%

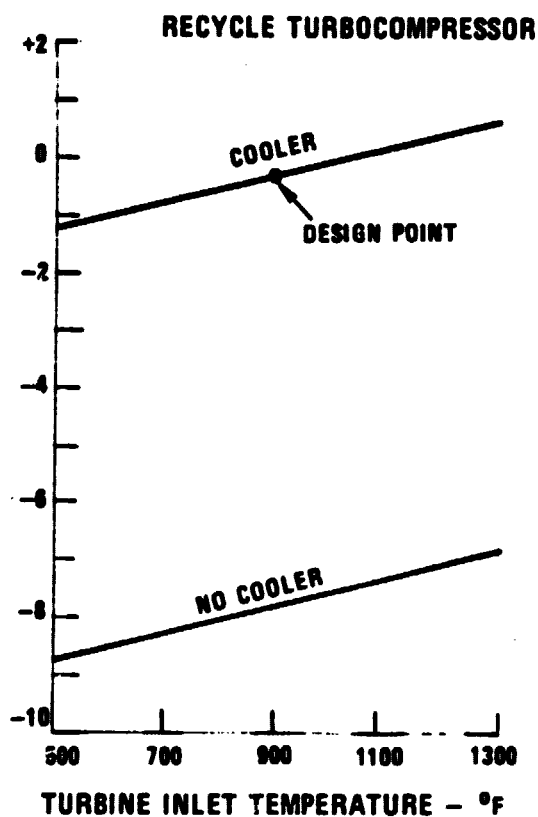
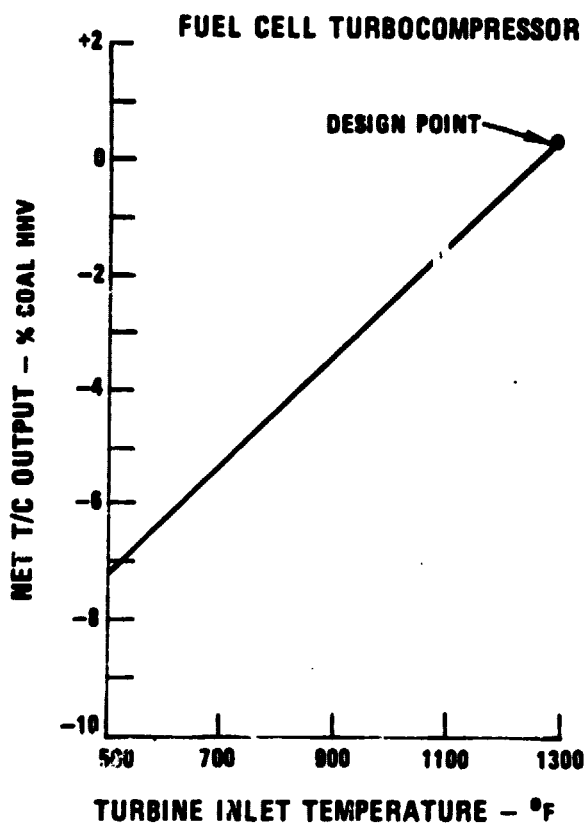


Figure 5-F. Impacts of Turbomachinery Operation for Basic Configuration

turbine preheat was also incorporated based on the results shown in the figure. In general, the maximum possible preheat temperature was found to be $\sim 900^{\circ}\text{F}$ based on system process heat availability. No process heat sources were available for higher preheating after allowances for the other thermal sink requirements were satisfied.

Optimum fuel cell pressure for operation of the single stage fuel cell turbocompressor is ~ 100 psia (Figure 5-G1). At lower pressures the pressure drops in the air loop reduce the turbine output relative to the compressor work. The shape of the curve at higher pressures is due to the assumption of a single stage machine and the limiting turbine inlet temperature of 1300°F . If compressor inter-cooling were used, the curve would tend to level off above 100 psia. It is desirable to use single stage machinery from a simplicity viewpoint, (fewer heat exchangers and easier thermal management). For a fixed design inlet temperature (1300°F), operation of the single stage turbocompressor at greater than ~ 100 psia implies an input of energy, in the form of sensible heat, to the system relative to the 100 psia point. In some cases this heat input may help provide overall thermal balance to the system.

Turbocompressor component efficiency has a large effect on net output of the machine (Figure 5-G2). Doubling of the machinery net output occurs between the nominal efficiency of 80%, used as the reference value in this study, and 90%. As with operating pressure, the component efficiency affects the thermal conditions of the process streams; lower efficiencies result in higher quality and quantity of process heat availability.

A map of resulting power plant efficiency is shown in Figure 5-H for the basic configuration. Of interest is the existence of a boundary for system thermal balance. Beyond the boundary line there is insufficient process heat (quality) availability. The basic configuration represents the first time this phenomenon has been observed, and is due to H_2O management in the recycle loop. As previously indicated in Figure 5-A, the thermal requirements for resaturation are significantly greater than previous studies with conventional gasifiers (no recycle). The re-

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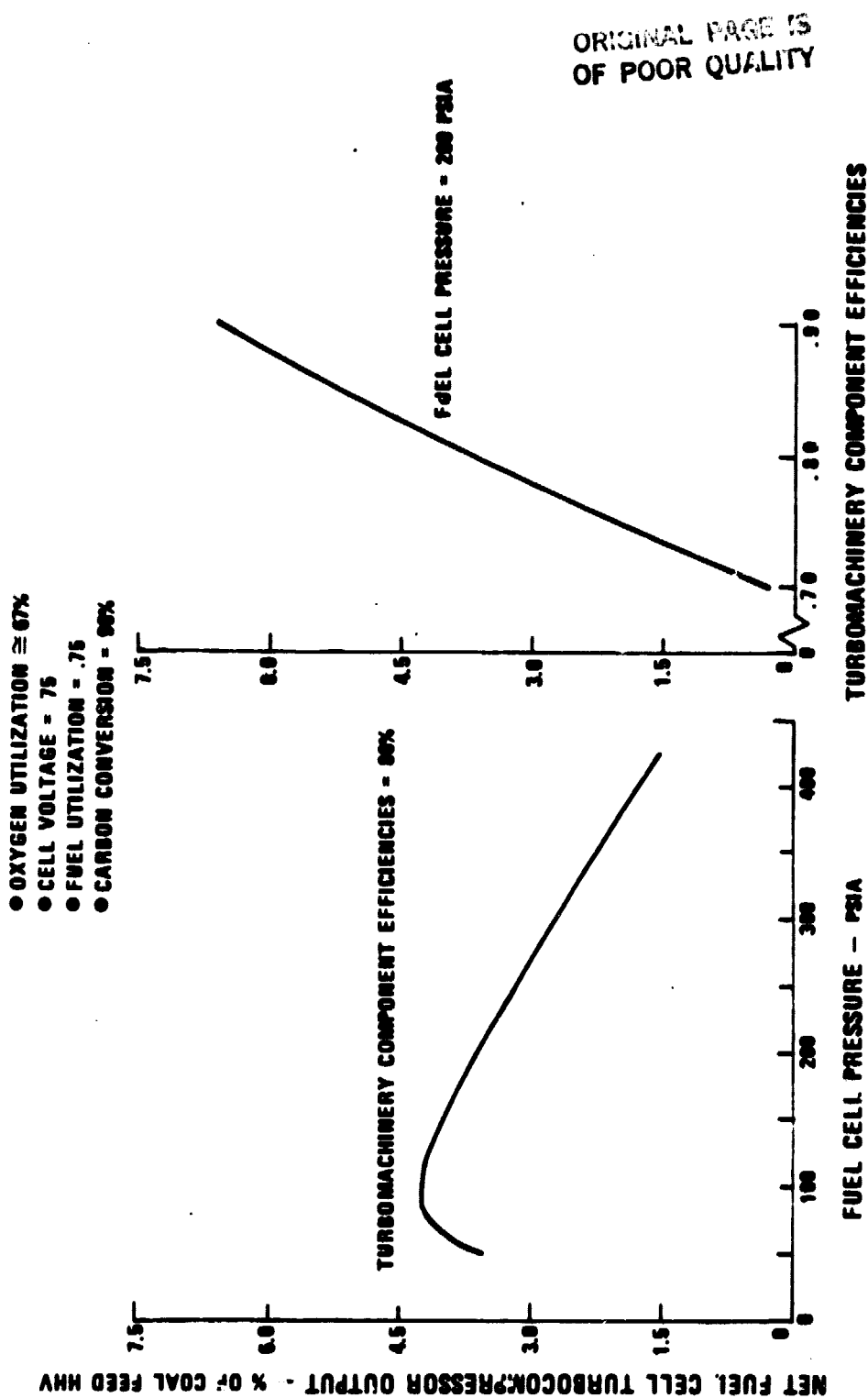


Figure 5-G. Fuel Cell Turbomachinery Performance for Basic Configuration

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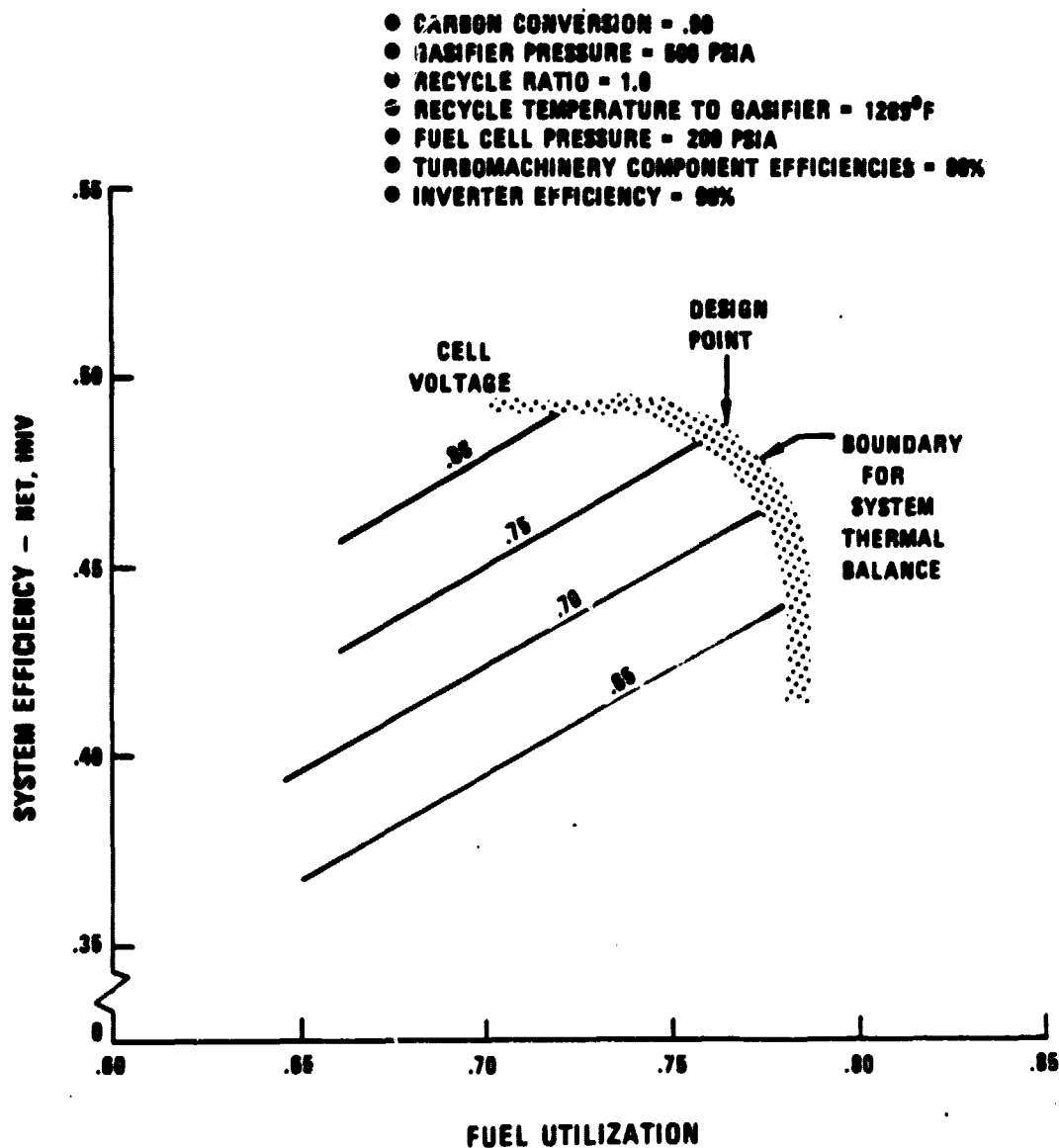


Figure 5-H. Map of System Efficiency vs. Fuel Cell Operation for Basic Configuration

cycle for the catalytic gasifier requires a high quantity of process heat for the latent heat of steam generation. A similar quantity of heat is available from the latent heat of condensation prior to low temperature desulfurization; however, this is low quality heat and except for some feedwater heating, is not usable as process heat and is dumped to the cooling towers. Thus, the system is constrained to operate at a relatively low fuel cell section efficiency to assume adequate process heat for thermal balance.

The design points selected for the thermodynamic design of the basic configuration include 75% fuel utilization at a cell voltage of 0.75. The process flow schematic and accompanying thermodynamic table are shown in Figure 5-1, and Table 5-A, respectively. The latter is a molar flow, based on a coal feed rate of one lb-mole of carbon feed to the gasifier per hour. Table 5-B summarizes the power plant operation and output.

Two process heat exchangers require special consideration. These two, located in the schematic prior to cathode inlet, transfer process heat from the cathode oxidant loop to the fuel side recycle loop. These are high temperature, air-to-fuel heat exchangers. Special design considerations may be required to ensure a leakproof design between the two fluid mediums. Any further effort on this system would require analysis in this area; however, the relatively low efficiency potential of this configuration as compared to previous studies precluded further study.

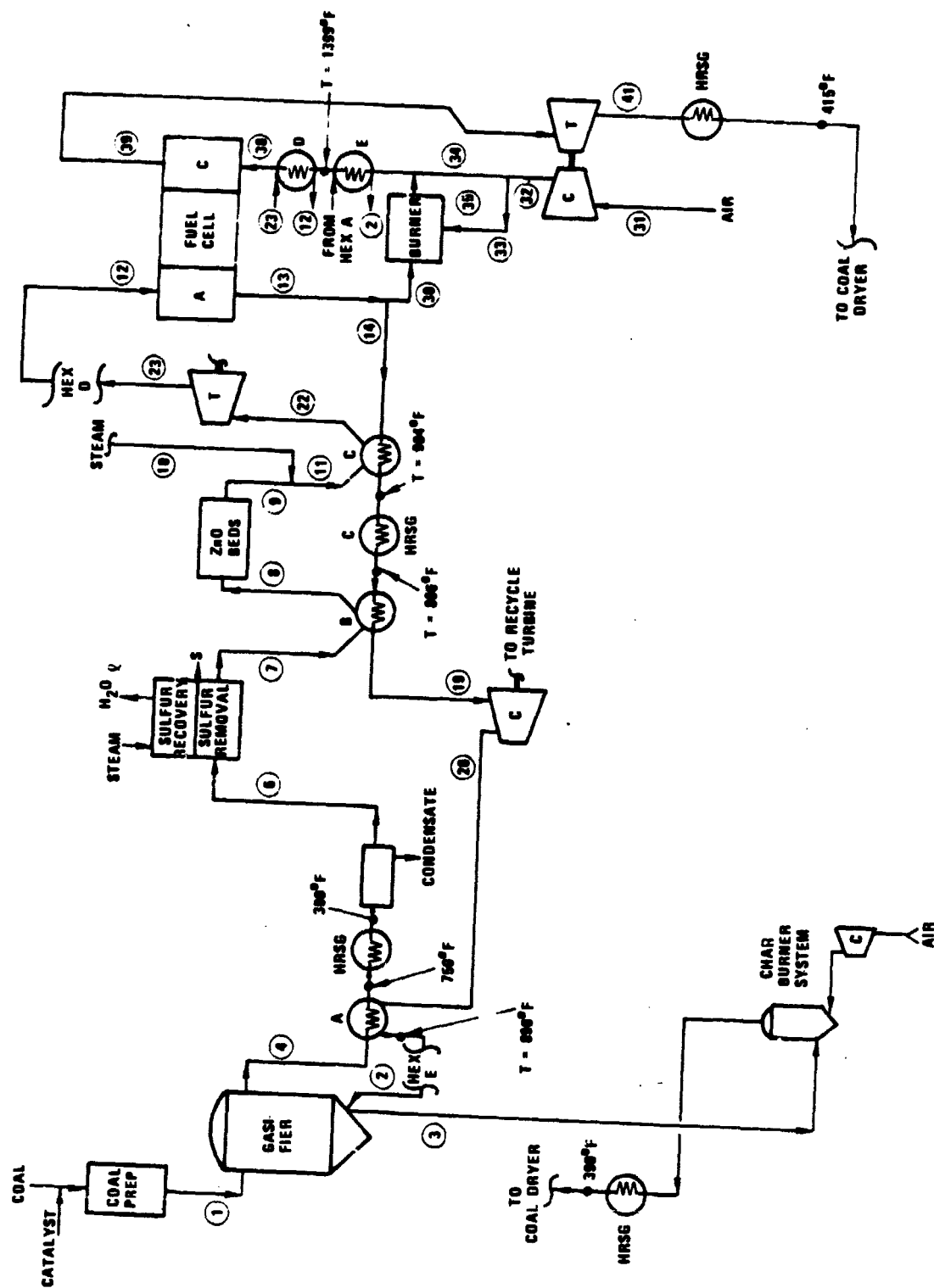


Figure 5-1. Basic Configuration Process Flow Schematic

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TABLE 5-A

BASIC CONFIGURATION THERMODYNAMIC TABLE

(Basis - 1 Lb-A/Hr Carbon Feed to Gasifier)

	TEMP	H2	H2O	CH4	CO	CO2	O2	N2	ENTH
1	400.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	-8155.
2	1200.	0.2701	1.8679	0.0014	0.2189	2.2892	0.0	0.0	-519926.
3	1097.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	45.
4	1097.	0.3159	1.2430	0.5089	0.2700	2.6306	0.0	0.0	-529106.
6	100.	0.3159	0.0071	0.5089	0.2700	2.6306	0.0	0.0	-457823.
7	70.	0.3159	0.0004	0.5089	0.2700	2.6306	0.0	0.0	-458122.
8	700.	0.3159	0.0004	0.5089	0.2700	2.6306	0.0	0.0	-434920.
9	700.	0.3159	0.0004	0.5089	0.2700	2.6306	0.0	0.0	-434920.
10	500.	0.0	2.9480	0.0	0.0	0.0	0.0	0.0	-280204.
11	623.	0.3159	2.9483	0.5089	0.2700	2.6306	0.0	0.0	-715124.
12	1190.	0.3159	2.9484	0.5089	0.2700	2.6306	0.0	0.0	-675206.
13	1300.	0.5401	3.7360	0.3028	0.4378	4.3784	0.0	0.0	-1029624.
19	312.	0.2701	1.8679	0.0014	0.2189	2.2892	0.0	0.0	-561482.
20	532.	0.2701	1.8679	0.0014	0.2189	2.2892	0.0	0.0	-551966.
21	0.	0.0	1.2432	0.0	0.0	0.0	0.0	0.0	0.
22	900.	0.3159	2.9483	0.5089	0.2700	2.6306	0.0	0.0	-696247.
23	734.	0.3159	2.9483	0.5089	0.2700	2.6306	0.0	0.0	-707723.
30	1300.	0.2701	1.8680	0.0014	0.2189	2.2892	0.0	0.0	-514812.
31	77.	0.0	0.0	0.0	0.0	0.0	1.5115	5.683	0.
32	791.	0.0	0.0	0.0	0.0	0.0	1.5115	5.683	63646.
33	791.	0.0	0.0	0.0	0.0	0.0	0.2983	1.121	12558.
34	791.	0.0	0.0	0.0	0.0	0.0	1.2133	4.562	51087.
35	2069.	0.0	2.1408	0.0	0.0	2.5095	0.0495	1.110	-302233.
37	1553.	0.0	2.1408	0.0	0.0	2.5095	1.2627	5.672	-451166.
38	1100.	0.0	2.1408	0.0	0.0	2.5095	1.2627	5.672	-500320.
39	1300.	0.0	2.1408	0.0	0.0	0.9000	0.4580	5.672	-246092.
41	863.	0.0	2.1408	0.0	0.0	0.9000	0.4580	5.672	-295313.
42	270.	0.0	2.1408	0.0	0.0	0.9000	0.4580	5.672	-323412.

*****CASE NO. 1. ENDS*****

TABLE 5-B

BASIC CONFIGURATION SUMMARY PERFORMANCE

Fuel Cell		
Cell Voltage		0.75 Volts
Fuel Utilization		0.75
Power Density		190 WSF
Pressure		200 PSIA
Gasifier		
Carbon Conversion		90%
Pressure		500 PSIA
Turbomachinery Efficiencies		
Turbines		0.80
Compressors		0.80
Power Plant Input/Output		
Coal In		220,597 Btu/Hr
Fuel Cell Out, DC		100,190 Btu/Hr
AC		98,186 Btu/Hr
Recycle Turbocompressor, AC		1,914 Btu/Hr
Fuel Cell Turbocompressor, AC		12,184 Btu/Hr
Parasite Power, AC		-4,676 Btu/Hr
Net Output, AC		107,608 Btu/Hr
Net Power Plant Efficiency		48.8%

6.0 ENHANCEMENT OPTIONS

Several configuration options were studied in an effort to increase system efficiency potential or to alleviate component design requirements (See Table 6-A). These enhancement options were designated according to the type of processes used in four subsystem areas:

- o Desulfurization
- o CO₂ management
- o Fuel gas saturation
- o Waste heat reforming

6.1 OPTION 1 - HIGH TEMPERATURE DESULFURIZATION

6.1.1 System Description

The major thermodynamic limitation of the basic configuration was the high thermal requirements for steam resaturation due to the condensation loss prior to low temperature desulfurization. The first configuration option was substitution of a high temperature primary desulfurization process for the low temperature Selexol[®] process used in the basic configuration. It was felt that this would alleviate H₂O management concerns and reduce system thermal balance constraints, resulting in reduced heat transfer requirements, higher fuel conversions (utilizations and cell voltages), and higher system efficiencies.

The primary desulfurization process assumed for this option is a packed bed of zinc-ferrite chemical absorbent pellets. The process is similar to that under development by the DOE at Morgantown Energy Technology Center. The beds operate cyclically between an absorbing and a regenerating mode. During absorption the beds can operate between 1000 and 1200°F at space velocities of 500-2000 hours⁻¹ and remove H₂S and COS down to levels between 2-8 ppmv. This permits use of zinc oxide (ZnO) secondary cleanup with operation similar to the basic configuration design, except that the fuel gas is wet. Equilibrium calculations

Table 6-A. Matrix of Configurations Studies

CONFIGURATION SUBSYSTEM	CONFIGURATION OPTION	ENHANCEMENT OPTIONS				4B
		1	2	3	4A	
DESULFURIZATION	BASIC					
	LOW TEMPERATURE	HIGH TEMPERATURE	LOW TEMPERATURE	HIGH TEMPERATURE		
CO ₂ MANAGEMENT	ANODE VENT TO BURNER AND CATHODE					
	BOILERS					
FUEL GAS SATURATION SCHEME FOR CARBON DEPOSITION PREVENTION						
	INTERNAL REFORMING					
WASTE HEAT RE- FORMING SCHEME						

↑ SAME AS BASIC CONFIGURATION

↑ SAME AS ENHANCEMENT OPTION 3

↑ SENSIBLE HEAT REFORMING

indicate that the ZnO beds are still capable of removing sulfur in the fuel gas to the levels required by the molten carbonate fuel cells.

For purposes of this study it was assumed that regeneration of the beds used only ambient air, although METC reported that steam dilution may be required to adequately control the temperature of the beds and prevent hot spots that fuse the absorbent. Sulfur absorbed on the bed is removed as SO_2 . Allowance for the compressor power for providing regeneration was included.

It was assumed that elemental sulfur was recovered using the Allied Chemical Corporation SO_2 Reduction process. In this process SO_2 is reduced with a fuel gas to obtain both elemental sulfur and H_2S ; the latter is reacted with unconverted SO_2 in the well-known Claus reaction. A tail gas incinerator converts all the remaining sulfur species to SO_2 prior to stack exhaust.

Although some development work has been done by Allied on the use of coal gas as a source for reducing fuel, methane was assumed in this study, based on conversations with Allied personnel. This is due to both the dilute nature of the SO_2 regeneration gas stream and the low Btu content of the recycle coal gas in this system. The methane used was assumed to equal 4% of the higher heating value of the coal feed.

The system schematic is shown in Figure 6-A and is similar to the basic configuration above, except that no raw gas cooling is done between the gasifier and the primary desulfurization process. Also, some process heat is available in cooling the fuel gas between the primary and secondary desulfurization processes.

6.1.2 Study Results

Comparison of the thermal requirements for steam resaturation (See Figure 6-B) shows that the thermal requirements are approximately halved for configuration Option 1 relative to the basic configuration. The requirements remain slightly greater, however, than those of systems using conventional type gasifiers without

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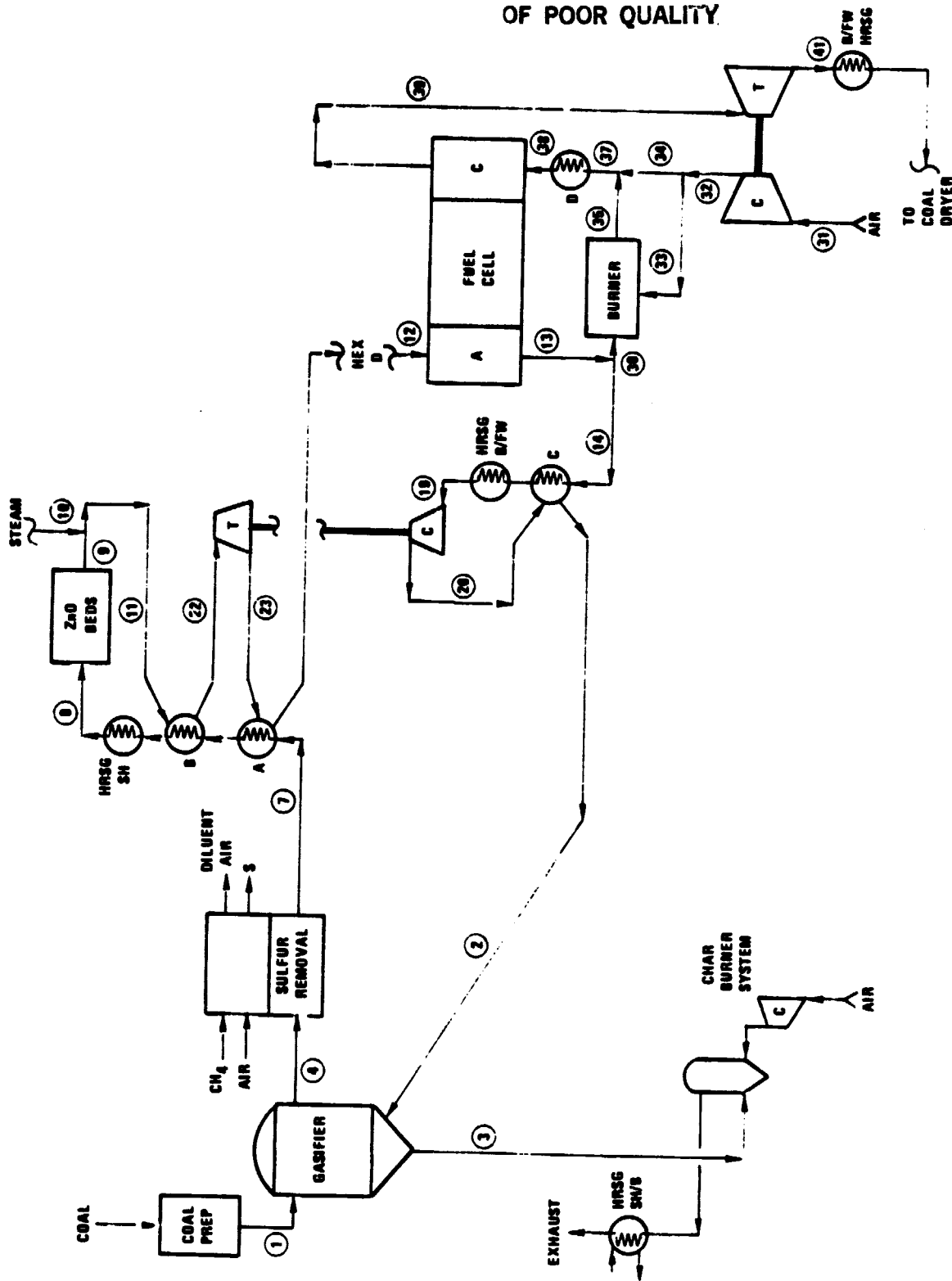


Figure 6-A. Configuration Option 1 System Schematic

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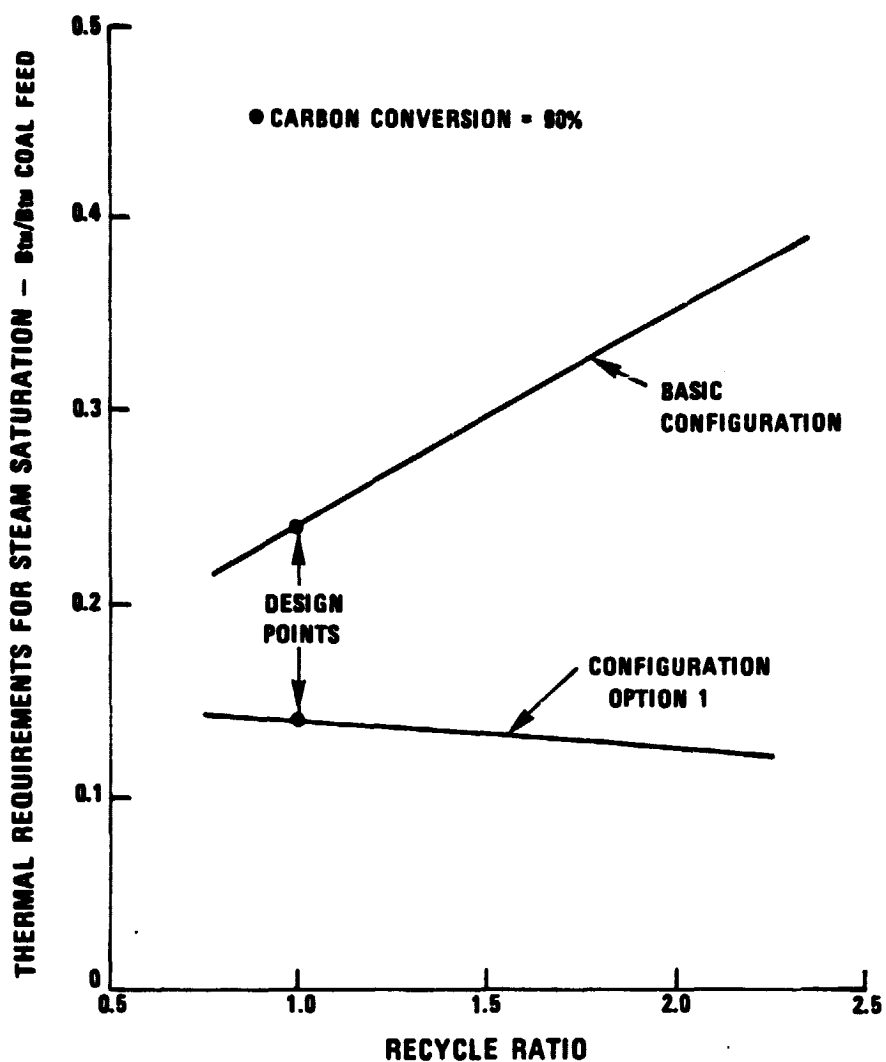


Figure 6-B. Comparison of Thermal Requirements for Steam Saturation

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TABLE 6-B

CONFIGURATION OPTION 1 THERMODYNAMIC TABLE
(Basis - 1 Lb-A/Hr Carbon Feed to Gasifier)

	TEMP	H2	H2O	CH4	CO	CO2	O2	N2	ENTH
1	400.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	-2155.
2	1200.	0.1602	1.9319	0.0002	0.1405	2.5620	0.0	0.0	-565553.
3	1062.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	45.
4	1062.	0.2731	1.2835	0.4834	0.2218	2.9975	0.0	0.0	-574671.
7	1064.	0.2814	1.2802	0.4834	0.2135	2.9057	0.0	0.0	-574671.
8	700.	0.2814	1.2802	0.4834	0.2135	2.9057	0.0	0.0	-595521.
9	700.	0.2814	1.2802	0.4834	0.2135	2.9057	0.0	0.0	-595521.
10	500.	0.0	1.6564	0.0	0.0	0.0	0.0	0.0	-157439.
11	600.	0.2814	2.9306	0.4834	0.2135	2.9057	0.0	0.0	-752960.
12	1100.	0.2814	2.9306	0.4834	0.2135	2.9057	0.0	0.0	-721062.
13	1000.	0.3204	3.3637	0.0003	0.2910	5.1239	0.0	0.0	-1120340.
14	1500.	0.1602	1.9319	0.0002	0.1405	2.5620	0.0	0.0	-560170.
19	512.	0.1602	1.9319	0.0002	0.1405	2.5620	0.0	0.0	-609205.
20	511.	0.1602	1.9319	0.0002	0.1405	2.5620	0.0	0.0	-600188.
22	900.	0.2814	2.7366	0.4834	0.2135	2.9057	0.0	0.0	-735391.
23	712.	0.2814	2.9306	0.4834	0.2135	2.9057	0.0	0.0	-749243.
30	1300.	0.1602	1.9319	0.0002	0.1405	2.5620	0.0	0.0	-560170.
31	77.	0.0	0.0	0.0	0.0	0.0	1.4823	5.573	0.
32	732.	0.0	0.0	0.0	0.0	0.0	1.4823	5.573	59350.
33	732.	0.0	0.0	0.0	0.0	0.0	0.1818	0.683	7273.
34	732.	0.0	0.0	0.0	0.0	0.0	1.3005	4.890	52073.
35	1705.	0.0	2.0924	0.0	0.0	2.7026	0.0301	0.676	-552392.
37	1333.	0.0	2.0924	0.0	0.0	2.7026	1.3307	5.566	-500319.
38	1100.	0.0	2.0924	0.0	0.0	2.7026	1.3307	5.566	-526112.
39	1301.	0.0	2.0924	0.0	0.0	0.9000	0.4294	5.566	-243535.
41	567.	0.0	2.0924	0.0	0.0	0.9000	0.4294	5.566	-295755.

*****CASE NO. 1-ENDS*****

the recycle. It should be noted that the composition of the recycle stream to the fuel cell, and therefore to the gasifier, is the same in Option 1 as the basic configuration (assuming similar fuel utilizations and operation of the gasifier).

Comparison of the system efficiency maps is shown in Figure 6-C for identical gasifier and recycle ratio conditions. The effect of the reduced thermal requirements is extension of the system thermal balance boundary to higher fuel cell efficiencies (product of cell voltage and fuel utilization) and a higher system efficiency potential. This represents the highest level of efficiency calculated in this study, and the highest known reported efficiency for an integrated coal gasifier-molten carbonate fuel cell power plant.

Further sensitivity studies with this configuration indicated an efficiency potential of up to 59% by using component turbomachinery efficiencies of 88% for the compressors and 92% for the turbines, as compared with the 80% values used in generating Figure 6-B.

Further comparison of Figure 6-C indicates that for the same fuel cell efficiency the calculated efficiency is slightly higher for the basic configuration than for Option 1. This is because of differences in the types and magnitudes of the parasitic requirements for primary desulfurization. Specifically, the Selexol[®] process in the basic configuration requires both electrical energy as well as significant steam requirements for stripping; the former is subtracted from the gross output to arrive at net system efficiency; whereas the latter adds to the total thermal requirements for steam generation in establishing the lower system thermal boundary limits. Alternatively, the major parasite for the high temperature desulfurization process is supplemental fuel, which detracts from net system efficiency. Thus, at a given fuel cell efficiency level within the permissible system thermal boundary, the effect of the electrical parasites for Selexol[®] is less than the effect of supplemental fuel, which accounts for the results in the figure.

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- CARBON CONVERSION = 90%
- GASIFIER PRESSURE = 500 PSIA
- RECYCLE RATIO = 1.0
- RECYCLE TEMPERATURE TO GASIFIER = 1200°F
- FUEL CELL PRESSURE = 200 PSIA
- TURBOMACHINERY COMPONENT EFFICIENCIES = 80%

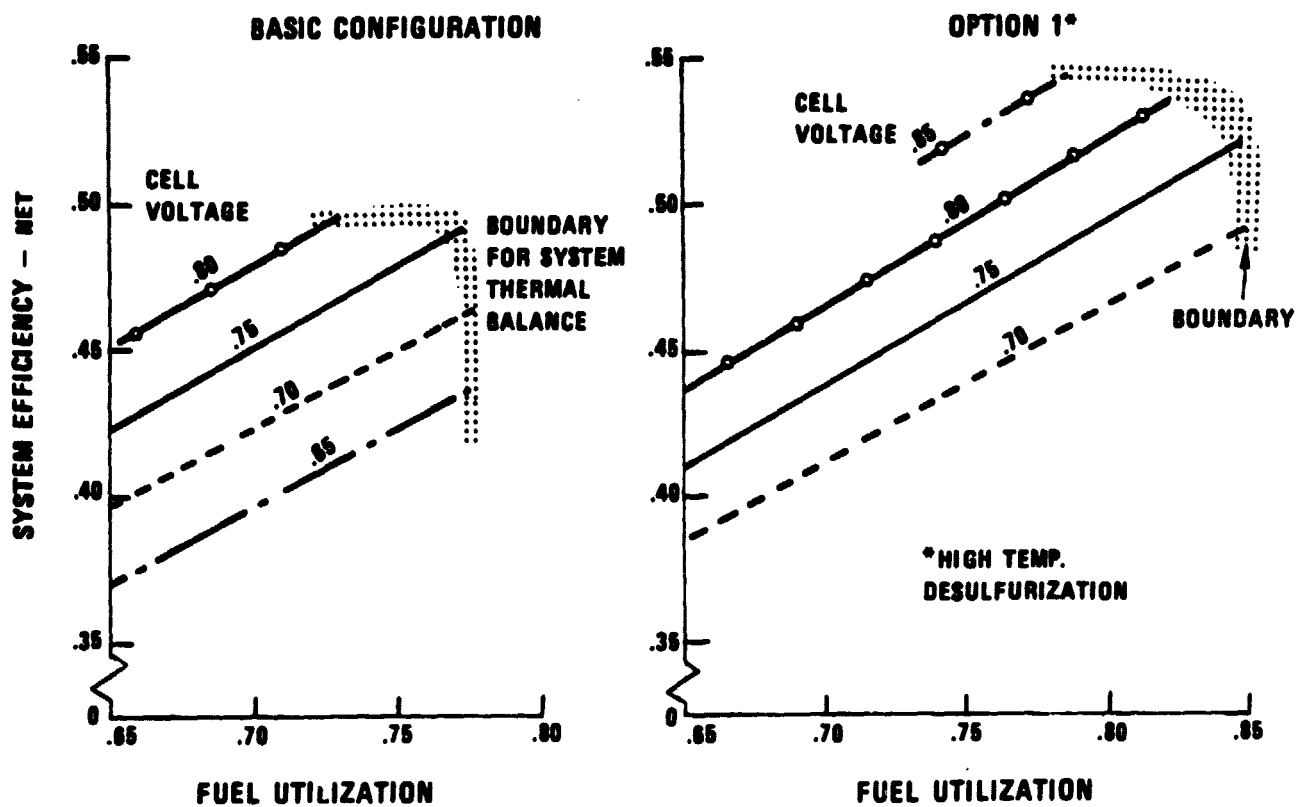


Figure 6-C. Comparison of System Efficiency Maps

Table 6-B shows the thermodynamics for this option at a design point corresponding to 59% net system efficiency. Table 6-C provides an overall summary of the system operation. The molar flows are based on an input coal feed of one lb-mole of carbon per hour.

This configuration option does not improve the low temperature operation required of the catalytic gasifier that was identified in the basic configuration. Also, high temperature air-to-fuel heat exchangers similar to the basic configuration are required. Finally, this configuration uses a non-commercial gas cleanup process. The three areas of concern for its development application to this system are:

- o Residual sulfur compounds and other potential contaminants that may not be absorbed.
- o Ability to control regeneration of the bed and prevent degradation of the absorbent.
- o Ability to recover elemental sulfur from the relatively dilute regenerated SO₂ gases.

6.2 OPTION 2 - CO₂ MANAGEMENT VIA PRODUCT CO₂ REMOVAL

6.2.1 System Description

A disadvantage of the two previous configurations was the resulting low temperature gasifier operation due to the high concentration of cell reaction product CO₂ in the recycle loop. Configuration Option 2 uses a CO₂ removal process in lieu of the anode recycle vent to provide CO₂ management. This significantly decreases the concentration of CO₂ in the recycle back to the gasifier, yielding recycle compositions similar to Exxon development experience.

The CO₂ removal process was assumed to be Purisol², a commercial physical solvent process licensed by American Lurgi. The CO₂ removal process was located in the anode recycle loop downstream of the fuel cells and prior to the recycle compressor. This placement maximizes the benefits to the gasifier operation by removal of CO₂ prior to feed to the gasifier. Also, this location represents the highest concentration of CO₂, which facilitates product CO₂ removal.

TABLE 6-C

CONFIGURATION OPTION 1 SUMMARY PERFORMANCE

Fuel Cell

Cell Voltage	0.78 Volts
Fuel Utilization	0.84
Power Density	144 WSF
Pressure	200 PSIA

Gasifier

Carbon Conversion	90%
Pressure	500 PSIA

Turbomachinery Efficiencies

Turbines	0.92
Compressors	0.88

Power Plant Input/Output

Coal In	220,597 Btu/Hr
Methane In	8,824
Fuel Cell Out, DC	116,701 Btu/Hr
AC	114,367 Btu/Hr
Recycle Turbocompressor, AC	4,240 Btu/Hr
Fuel Cell Turbocompressor, AC	21,771 Btu/Hr
Parasite Power, AC	-3,276 Btu/Hr
Net Output, AC	137,102 Btu/Hr
Net Power Plant Efficiency	59.8%

Since only bulk removal of acid gas was needed, the Purisol process was assumed to operate in a pressure swing absorption mode with no thermal requirements. The process operate at near ambient (100°F) temperature. It was further assumed that the process operated in order to maintain carbon balance across the loop; i.e., the CO₂ removed was equal to the product CO₂ from the anode reaction as well as the equivalent carbon entering the loop from the gasifier reaction. If the water management is also properly controlled, the recycle can operate without need of a vent stream, with only CO₂ and H₂O being removed in such a way that mass balance is achieved. This permits unity fuel utilization in the fuel cell, which offers the potential of higher system efficiency.

Initially the H₂O management approach was assumed to be similar to the basic configuration. The H₂O management problems encountered in the basic configuration, however, were doubled here due to the drying of the streams via condensation in two areas of the recycle loop; namely, the desulfurization process and the CO₂ removal process. The thermal requirements for steam generation to resaturate the recycle downstream of both processes were extreme and did not permit a system thermal balance at any reasonable level of system efficiency. A "novel" resaturation scheme, therefore, was employed as part of this configuration.

The basis of the scheme was to use the low grade heat of condensation to provide the majority of the heat of vaporization for resaturation. This prevented the excessive loss of low grade system process heat to the cooling towers. The scheme, shown in Figure 6-D, consists of two packed beds, or water tray towers, one operating as a contact saturator, the other, as a contact cooler. The dry gas to be wetted enters the saturator in counter-current flow to a warmer water stream. The gas is saturated to a temperature approximating the temperature of the incoming water. Unused water is recycled back to the contact cooler; some cooling of the water may be provided. This cooled water stream is contacted against the wet gas stream, cooling and drying it via condensation to a saturation temperature approximating the cool water inlet temperature. Some heat is added to the water loop prior to recycle to the contact saturator.

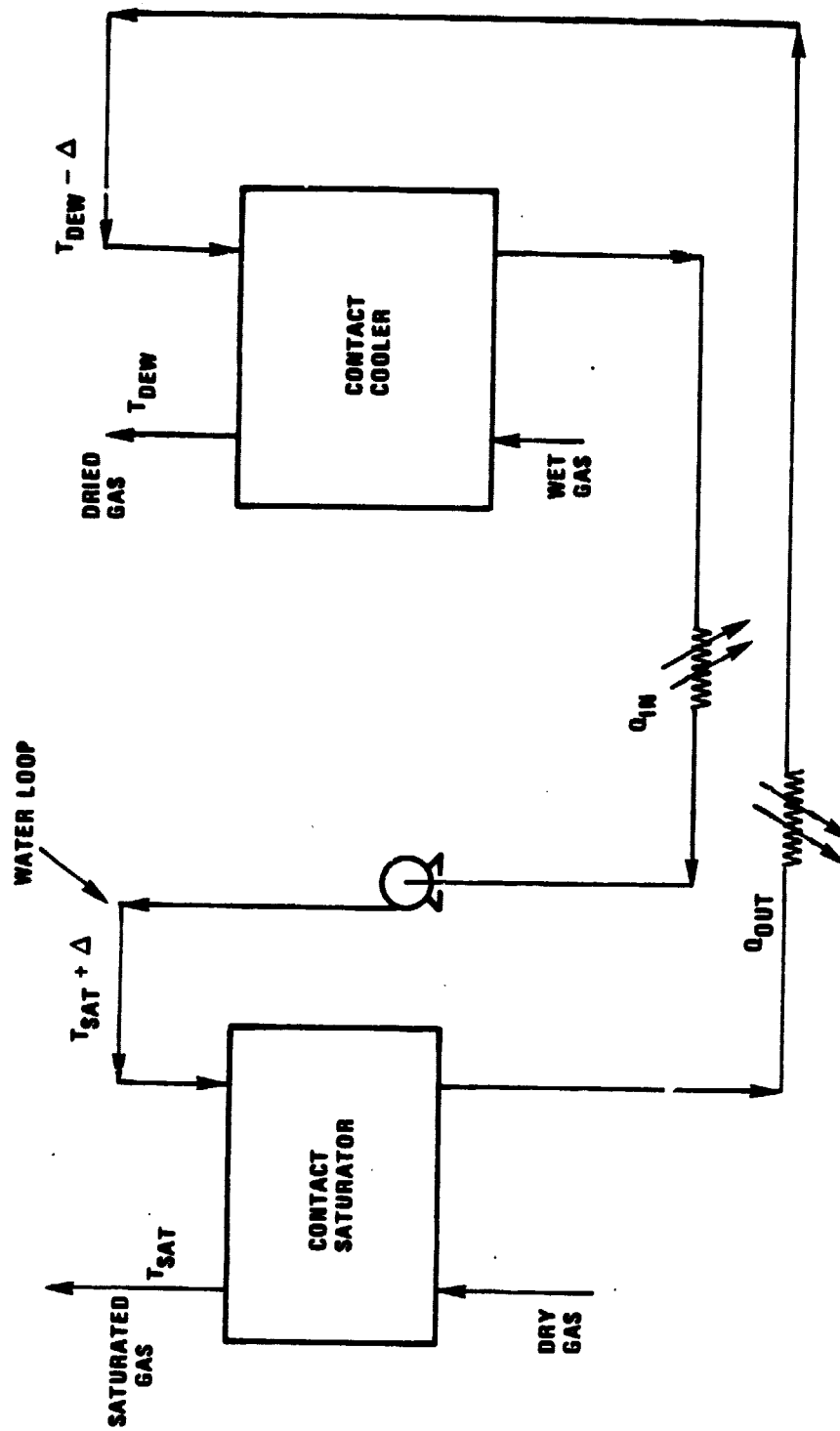


Figure 6-D. Novel H₂O Resaturation Scheme for Option 2

Incorporation of this H₂O management scheme into the system is shown in Figure 6-E. Two sets of contact coolers/saturators were used to minimize steam generation via conventional process heat recovery boilers.

6.2.2 Study Results

Comparison of the thermal requirements for steam generation between this configuration option and the two previous configurations is shown in Table 6-D. Use of the contact cooler/saturators significantly reduces the thermal requirements of Option 2 to a level comparable to Option 1.

Figure 6-F shows the impact of CO₂ management via the CO₂ removal process on gasifier operation. The combined concentration of synthesis gas is 52% vs. approximately 57% based on Exxon development experience. The effect of the richer synthesis gas recycle to the gasifier of this option relative to the previous options is to increase the temperature of gasifier operation to the level of Exxon experience. This is illustrated in the lower bar graph of the figure.

The resulting efficiency for this configuration was less than might be expected for unity fuel utilization. This can be understood by reference to Figure 6-G, which compares the net gross a.c. outputs of the fuel cell and turbomachinery for the Option 2 configuration to the basic configuration. For the basic configuration, the turbomachinery provides generating capacity, whereas, for Option 2, additional shaft power must be provided. This is primarily due to the need for recompression of the separated CO₂ prior to feed to the fuel cell cathode. Secondly, the inlet temperatures to both the fuel cell and recycle turbine are lower than previous configurations in order to match both thermal quality and quantity for process heat recovery. This is indicated in the schematic of this configuration, Figure 6-E, which shows that no recycle let down turbine preheat is used and that some heat is removed from the cathode exhaust prior to inlet to the fuel cell expansion turbine.

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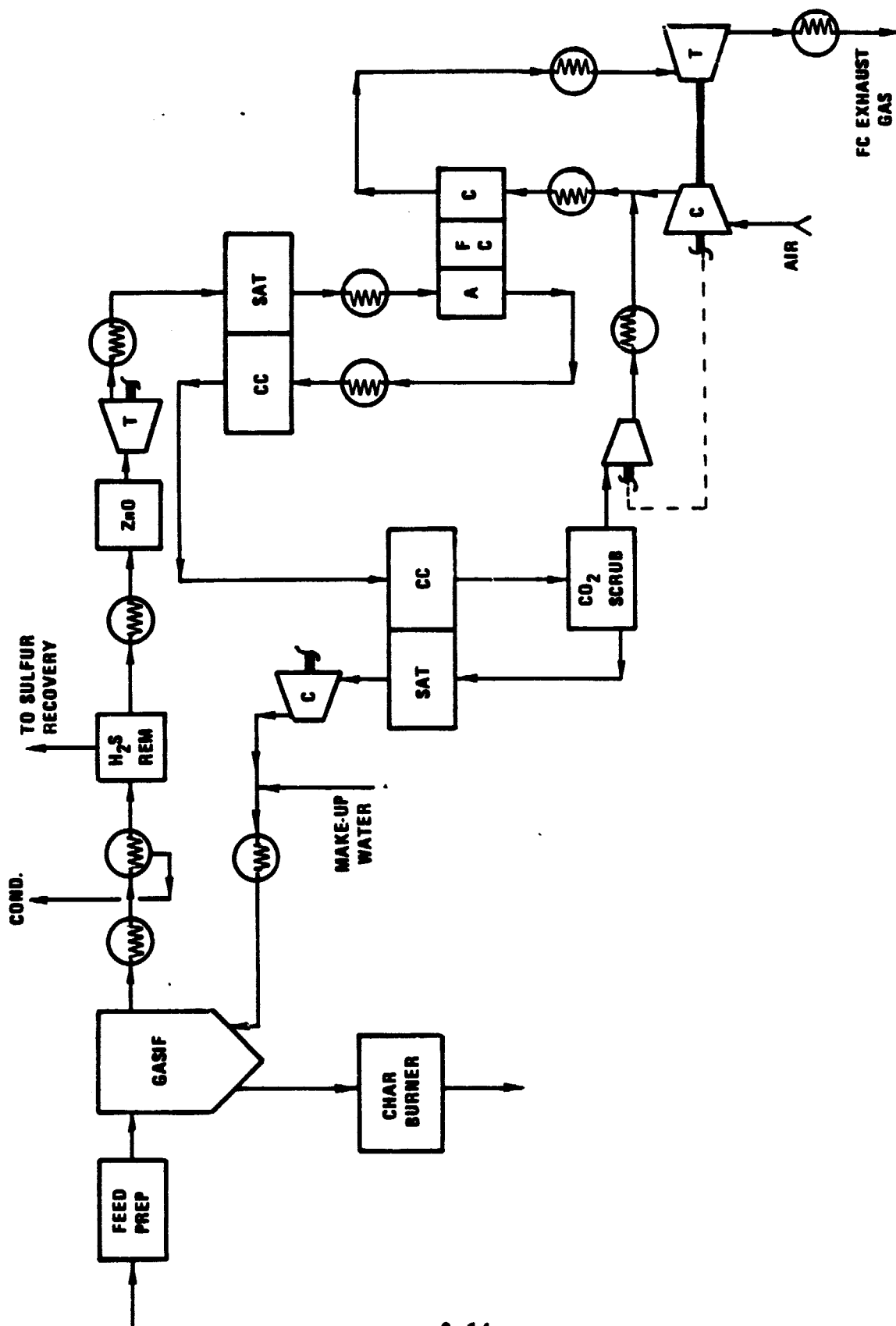
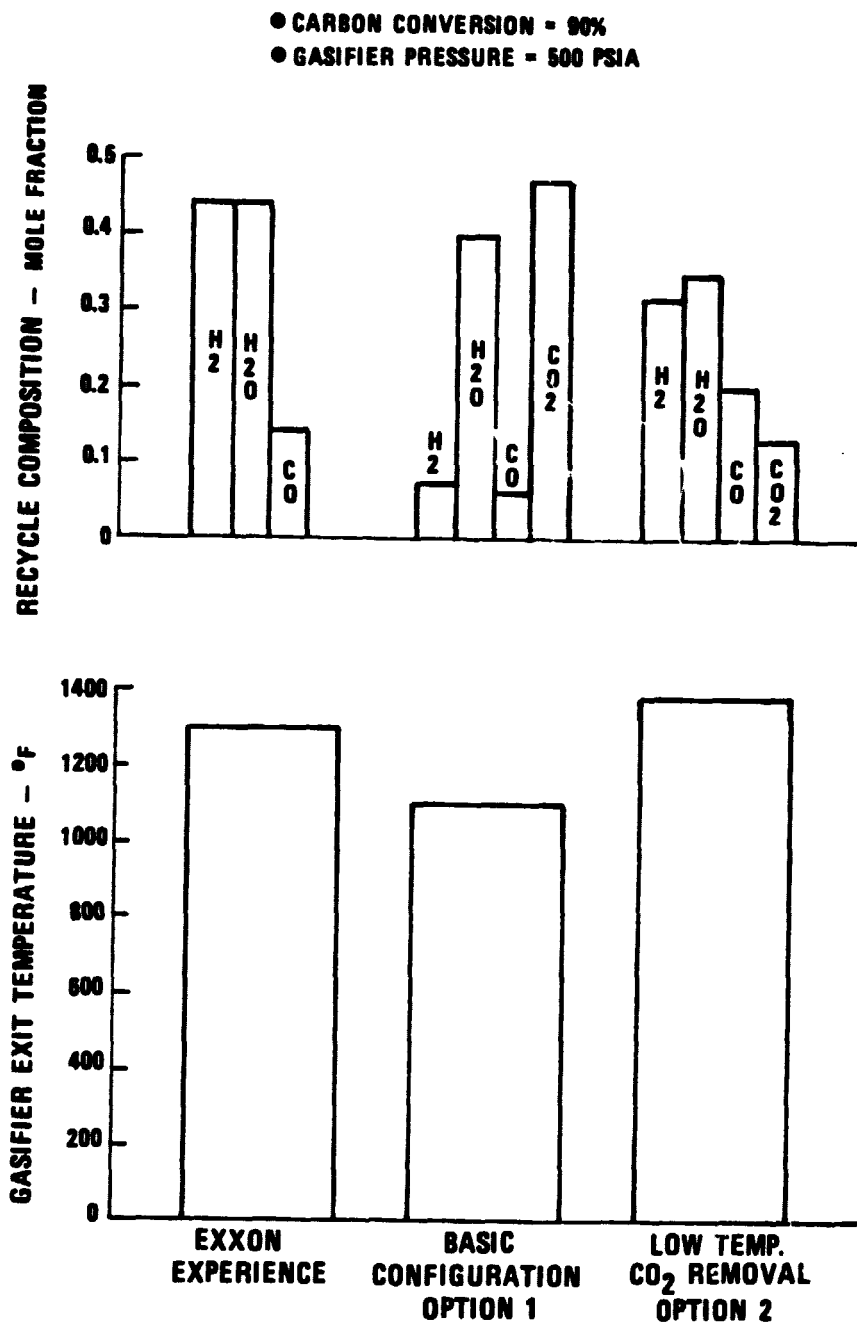


Figure 6-E. Configuration Option 2 System Schematic

TABLE 6-D

COMPARISON OF THERMAL REQUIREMENTS FOR STEAM SATURATION

	Basic Configuration	Option 1	<u>Option 2</u>	
			Without Novel Scheme	With Novel Scheme
Thermal Requirements for Saturation of Recycle Loop (~ Btu/Btu Coal Feed)	0.24	0.14	0.31	0.18

Figure 6-F. Impact of CO₂ Management on Gasifier Operation

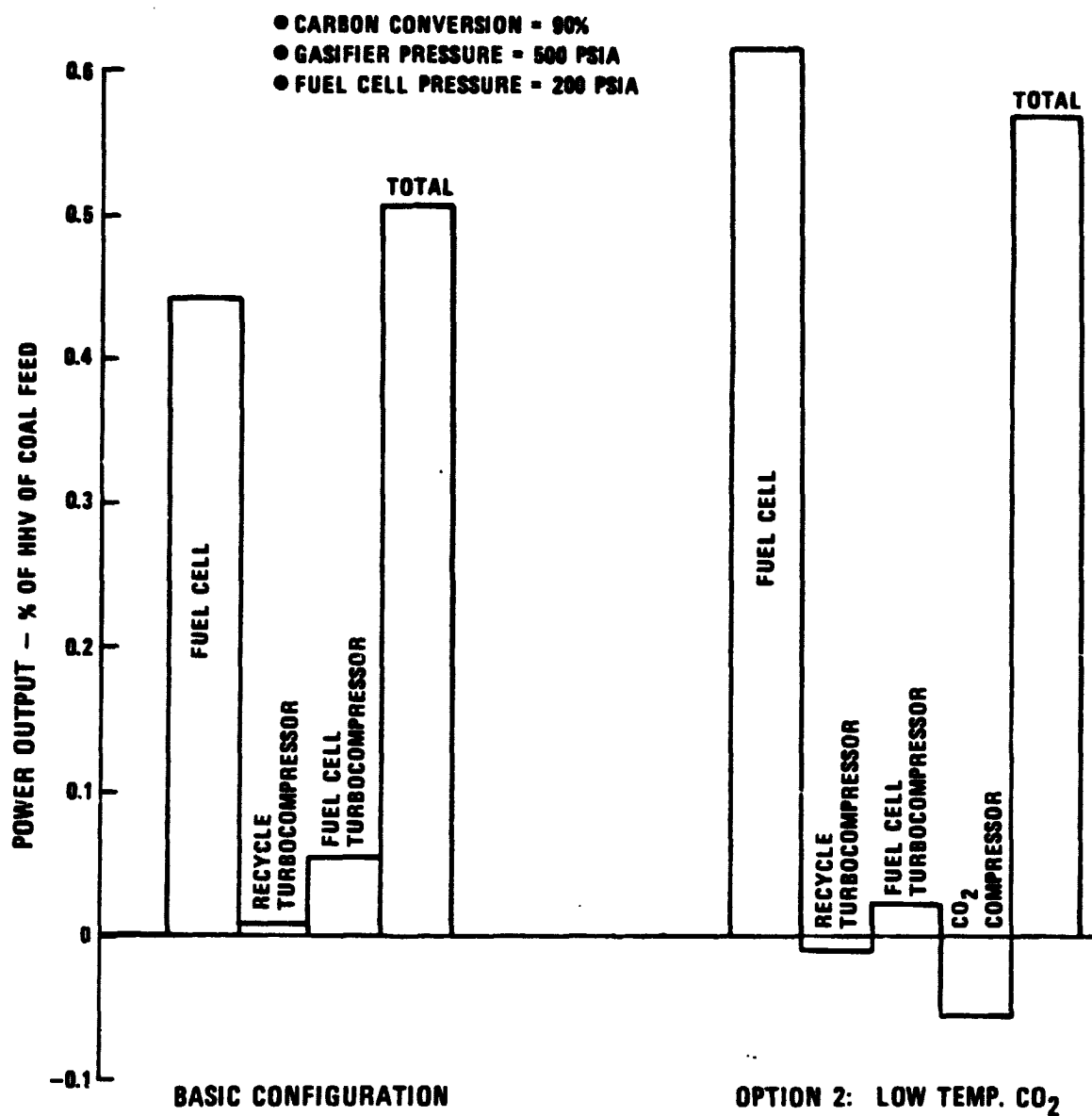
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Figure 6-G. Power Output Comparison

The efficiency potential of this system configuration is 51.5% (see Table 6-E). As indicated from the previous description, this configuration is based on use of commercial processes in the recycle loop, and results in gasifier operating conditions in the range of development experience. The use of contact coolers/saturators is not expected to be a development requirement. Furthermore, the use of this H₂O management scheme eliminates the need for high temperature air-to-fuel heat exchangers in the system design. The high efficiency potential coupled with these other system advantages make this an attractive configuration option for further study.

6.3 OPTION 3 - ADVANCED PROCESSES FOR DESULFURIZATION AND CO₂ REMOVAL

6.3.1 System Description

The previous Option 2 configuration used low temperature commercial processes for both CO₂ removal and desulfurization and required a novel H₂O management scheme to attain its high efficiency potential. In an attempt to simplify the anode recycle loop and further reduce thermal requirements for steam saturation, it was assumed that high temperature processes would be substituted for the low temperature Selexol[®] and Purisol[®] processes used in Option 2. This offered the possibility of increased efficiency relative to Option 2, while maintaining two advantages of Option 2: improved gasifier operation and unity fuel utilization.

The desulfurization process assumed in Option 3 is the packed bed zinc-ferrite, absorbent type used in Option 1, with identical operation. The process for CO₂ removal was not specified. It was assumed to be selective, operating in a pressure swing mode and operating at temperatures in excess of 350-400°F in order to avoid condensation. The process could use either a chemical or physical solvent or, more likely, a solid absorbent or membrane. Thus, similarly to the operation of Purisol previously referred to, no thermal was required, but the separated CO₂ had to be recompressed to fuel cell pressure.

TABLE 6-E

SUMMARY OF DESIGN CHARACTERISTICS FOR CONFIGURATION OPTION 2

Product CO ₂ Management	Low Temperature Removal
Desulfurization	Low Temperature Physical Solvent
H ₂ O Saturation Scheme	Novel Contact Saturation
Waste Heat Reforming	Internal Reforming
Air to Fuel High Temperature Heat Exchangers	Yes
Fuel Cell: Cell Voltage	0.82 Volts
Fuel Utilization	1.0
Cell Pressure	200 PSIA
Cell Power Density	197
Gasifier: Carbon Conversion	90%
Operating Pressure	500 PSIA
Operating Temperature Range	1300 - 1400°F
Overall Coal Pile to Busbar Efficiency	51.5%

6.3.2 Study Results

The results for Option 3 were not encouraging relative to the high degree of development required. The efficiency potential, (See Table 6-F) was approximately 54%, slightly higher than that of Option 2. No steam generation was required for resaturation of the anode loop and no air-to-fuel heat exchangers are required. The resulting efficiency potential with use of both a non-commercial desulfurization process and an unknown CO₂ removal process, did not provide an incentive for further study of this configuration option.

6.4 OPTIONS 4A AND 4B - SENSIBLE HEAT REFORMING

6.4.1 System Description

Two configurations were studied that assumed use of sensible heat reforming (SHR) of the fuel gas from the gasifier in lieu of internal reforming. Use of SHR eliminates design and development of an internal reforming fuel cell stack. Physical separation of the reforming and cell reaction processes also provides maintenance flexibility between the reform catalyst and the anode itself.

In order to focus this area of study, SHR was substituted for internal reforming in two previously discussed configurations; namely, the basic configuration, and Option 3, using advanced processes for desulfurization and CO₂ removal. It was felt that comparison of these two configurations using SHR and internal reforming would indicate the range of impacts that could be expected in this system concept.

The basic configuration with SHR is identified as Option 4A. The SHR requires sufficient flow of gases through the bed to provide the endothermic heat of reforming via sensible heat from the gas stream. The quantity of sensible heat available from the gas per unit volume flow is limited by the inlet temperature and allowable temperature drop across the beds. It was determined that the gasifier - fuel cell recycle flow would have to be greatly increased relative to the reference conditions (recycle ratio of 1) in order to sustain the SHR. As previously dis-

TABLE 6-F

SUMMARY OF DESIGN CHARACTERISTICS FOR CONFIGURATION OPTION 3

Product CO ₂ Management	High Temperature Pressure Swing Removal
Desulfurization	High Temperature Absorption
H ₂ O Saturation Scheme	None
Waste Heat Reforming	Internal Reforming
Air to Fuel High Temperature Heat Exchangers	No
Fuel Cell:	
Cell Voltage	0.82 Volts
Fuel Utilization	1.0
Cell Pressure	200 PSIA
Cell Power Density	180
Gasifier:	
Carbon Conversion	90%
Operating Pressure	500 PSIA
Operating Temperature Range	1300 - 1400°F
Overall Coal-Pile-to-Busbar Efficiency	54.1%

cussed in Section 5, this would significantly detract from system efficiency. Therefore, a second recycle loop was included around the fuel cell anode to transfer the cell reaction waste heat to the SHR. Similarly, the same approach is used in substituting SHR into Configuration Option 3; the resulting configuration is referred to as Option 4B. The use of a second recycle adds control complexity to the system and requires use of a high temperature recycle compressor.

6.4.2 Study Results

The major impacts of using SHR in lieu of internal reforming are in the area of cell performance and system efficiency. Figure 6-H compares the cell power density projections between SHR Options 4A and 4B, and the analogous systems, the basic configuration, and Option 1, respectively. Little performance loss is projected for substitution of SHR in the basic configuration. This is due to operation at low cell efficiencies which minimizes dilution of reactant fuel gases. At higher cell efficiencies, which is the case with the options using a CO₂ removal process resulting in unity fuel utilizations, the fuel gases in the cell become more dilute. This is accentuated by incorporation of SHR, as in Option 4B, for three reasons:

- o The recycle of the SHR dilutes the average concentration of fuel gas across the cell.
- o The reform equilibrium temperature at the exit of the SHR bed tends to be lower than at the exit of an internal reforming cell anode, resulting in less reforming and less available synthesis gas to the cell.
- o The products of reforming remain throughout the SHR bed, which tends to limit reforming. In the internal reforming cell the products are continually consumed in the cell reaction.

As shown in Figure 6-H the cell power density is reduced by approximately 40% between Option 3 and the analogous configuration with SHR (Option 4B).

The system design characteristics for the SHR options are shown in Table 6-G. These results assume that the SHR systems operate at the same cell efficiency as

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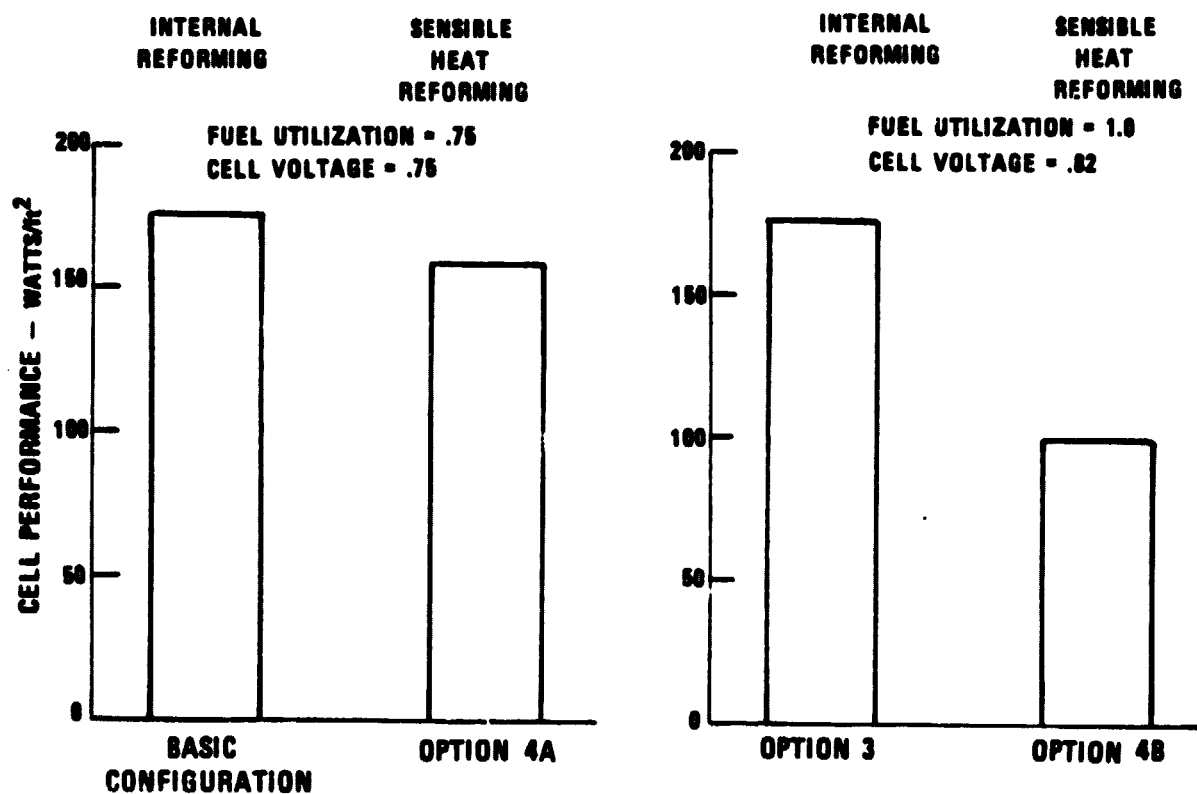


Figure 6-H. Cell Performance Comparison

TABLE 6-G

SUMMARY OF DESIGN CHARACTERISTICS FOR
CONFIGURATION OPTIONS 4A AND 4B

	Option 4A	Option 4B
Product CO ₂ Management	Vent	High Temperature Pressure Swing Removal
Desulfurization	Low Temperature Physical Solvent	High Temperature Absorption
H ₂ O Saturation Scheme	Steam Injection	None
Waste Heat Reforming	← Sensible Heat Reforming →	
Air to Fuel High Temperature Heat Exchangers	Yes	No
Fuel Cell:		
Cell Voltage	0.75	0.82
Fuel Utilization	0.75	1.0
Cell Pressure	200 PSIA	200 PSIA
Cell Power Density	160 WSF	100 WSF
Gasifier:		
Carbon Conversion	← 90% →	
Operating Pressure	← 500 PSIA →	
Operating Temperature Range	1050 - 1100°F	1300 - 1400°F
Overall Coal Pile to Busbar Efficiency	46.7%	52%

their analogous internal reforming options, which was the basis of the cell performance comparison in the preceding figure. The loss in efficiency for the SHR option is due to the parasite power for the second recycle loop and is estimated to result in approximately a two point loss in system efficiency.

7.0 CONCLUSIONS AND RECOMMENDATIONS

The concept of integrating a catalytic coal gasifier with a waste heat reforming molten carbonate fuel cell power section has several attractive features:

- o Elimination of the need for an O_2 plant for gasification.
- o Elimination of a steam turbine bottoming cycle.
- o Use of a waste heat reforming molten carbonate fuel cell power section operating at conditions consistent with the present direction of fuel cell technology development at UTC.
- o High power plant efficiency potentials in the range of 51.5-59%, depending on the degree of process development required. Comparison of the efficiency potentials for the range of configurations studied is shown in Figure 7-A.

The elimination of an O_2 plant for gasification is an economic benefit and potentially an operational benefit. Typically, studies have shown that O_2 plant costs account for 25-35% of the total costs of gasification. Furthermore, there are concerns over both the turndown range and response rate for O_2 plants integrated with coal gasifiers for a central power station mode of operation.

Elimination of the steam bottoming cycle similarly has economic and operational benefits, as well as the advantage of reduced natural resource requirements. Typically, in the conventional conceptual design of an integrated gasifier - fuel cell power plant, the steam turbine and associated auxiliaries account for 20-25% of the power section costs. Without the steam turbine, more than 90% of the total power plant output is associated with the electrochemical fuel cell reaction and solid state inverters. Such a power plant would have several of the features of the smaller dispersed utility fuel cell power plants presently under development by UTC, including faster response potential and increased utility stability, an example of which is VAR control. Finally, lack of a steam turbine cycle significantly reduces power plant H_2O makeup requirements associated with heat rejection through the cooling towers. A comparison of these makeup water requirements is shown in Figure 7-B.

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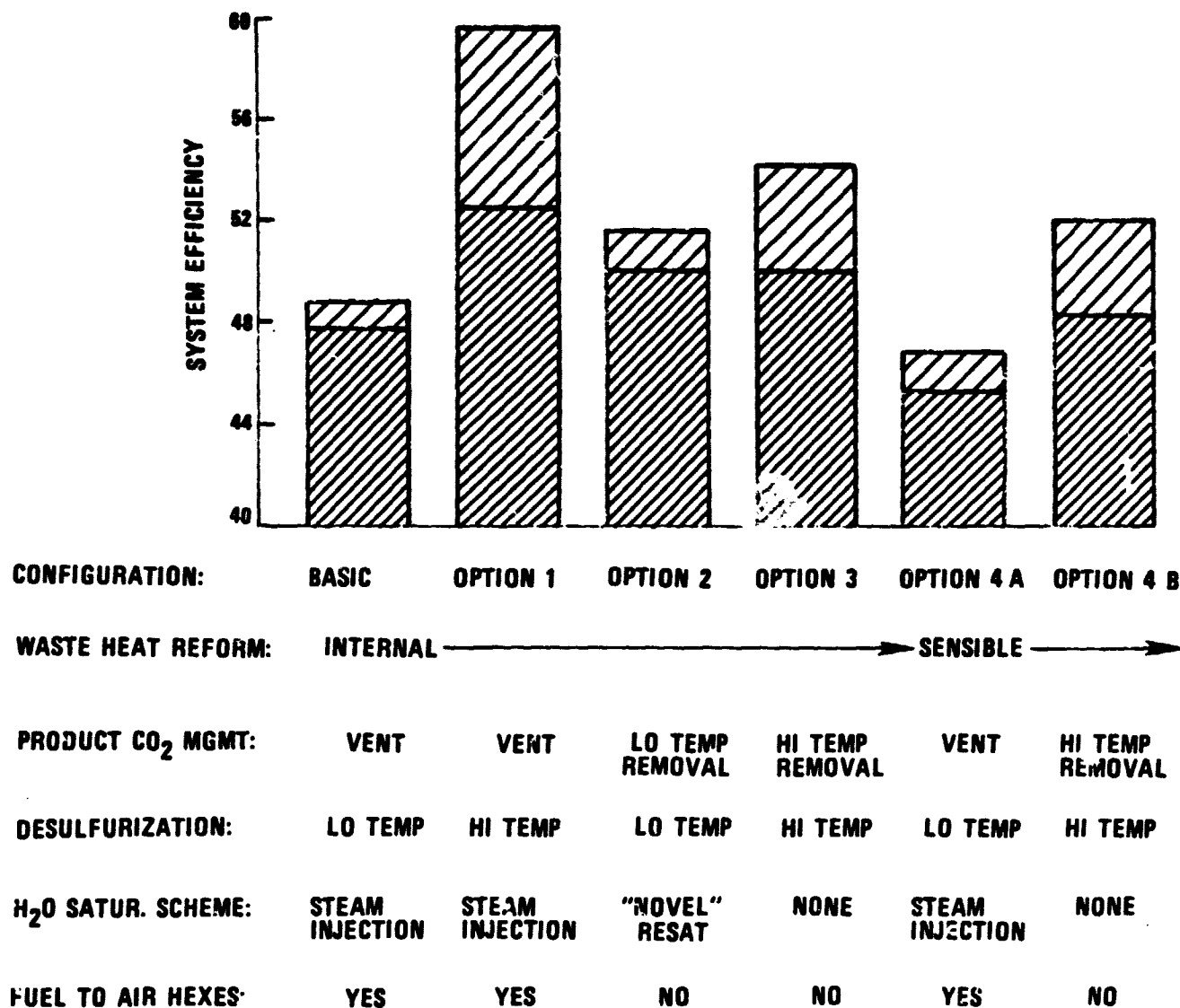


Figure 7-A. System Efficiency Comparison

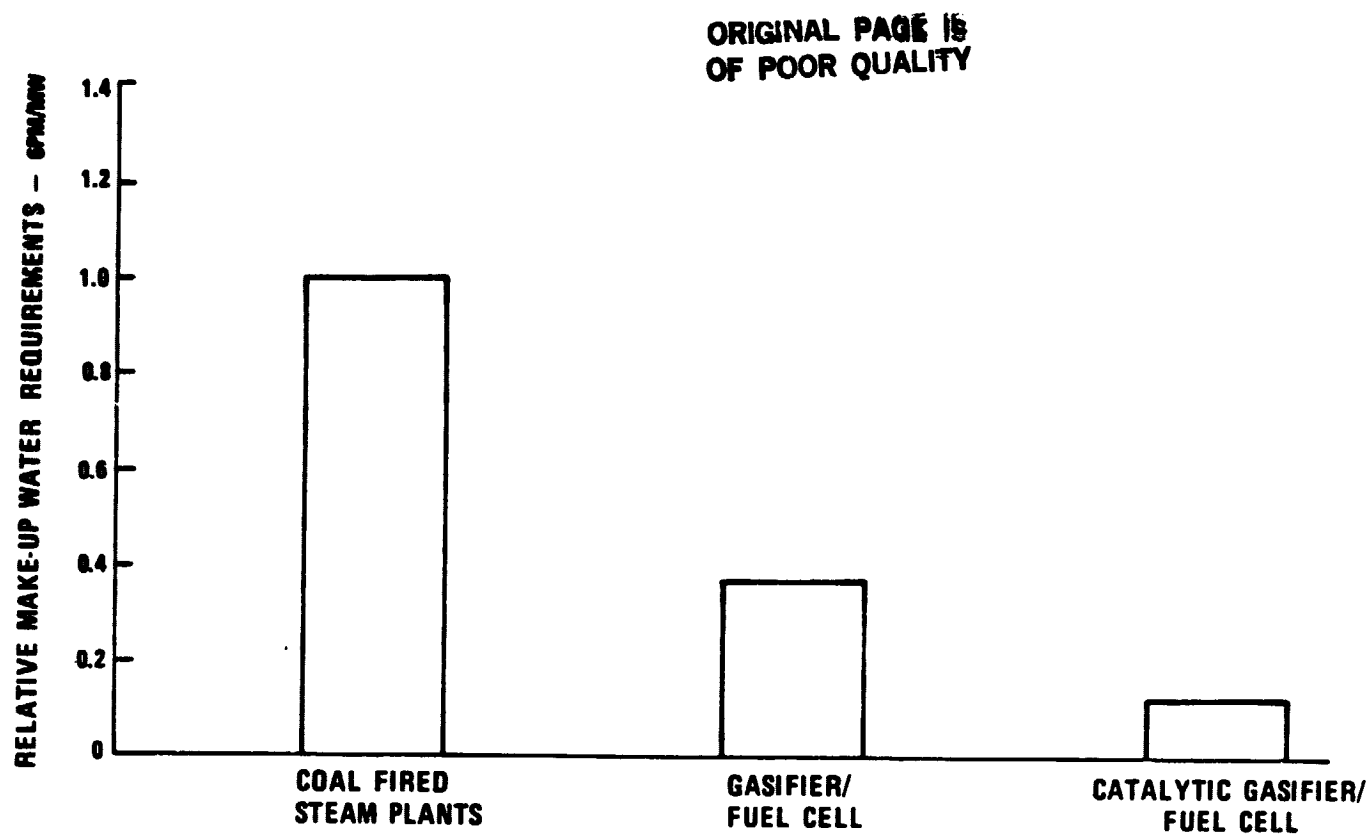


Figure 7-B. Comparison of Water Requirements

The studies showed that the assumed operating conditions for the fuel cell with either of the waste heat reforming approaches were consistent with present technology development directions. Operation of the fuel cell at pressures greater than 200 psia was not beneficial to the system.

Operation of the cell at higher temperatures is also not considered beneficial. Although this could result in higher recycle temperatures to the gasifier, the sensitivity studies for the basic configuration indicated that this temperature increase would have to be significant ($\sim 400^{\circ}\text{F}$) in order to bring the gasifier temperature into the range of Exxon development experience. This would also be at the expense of added process heat recovery equipment.

Comparison of the efficiency potentials for the integrated catalytic gasifier - molten carbonate fuel cell power plant concept with previously studied conventional gasifier - fuel cell power plants is shown in Figure 7-C. Increased efficiencies are possible relative to the conventional systems. However, additional development requirements are needed, and the degree of development increases with increased efficiency potential. Table 7-A lists the development requirements in order of increased system efficiency potential.

In the area of catalytic gasifiers, heat recovery is important; therefore, development of char burners should be part of any catalytic gasifier development program.

The system configuration using commercial processes for desulfurization and CO_2 removal (Option 2) is significant in that it yields as high efficiency as the conventional systems and does not require process development other than the gasifier and fuel cells, with the former operating at Exxon development conditions. Further study of this system is recommended, and should include an economic assessment.

It is further recommended from this study to follow two areas of development:

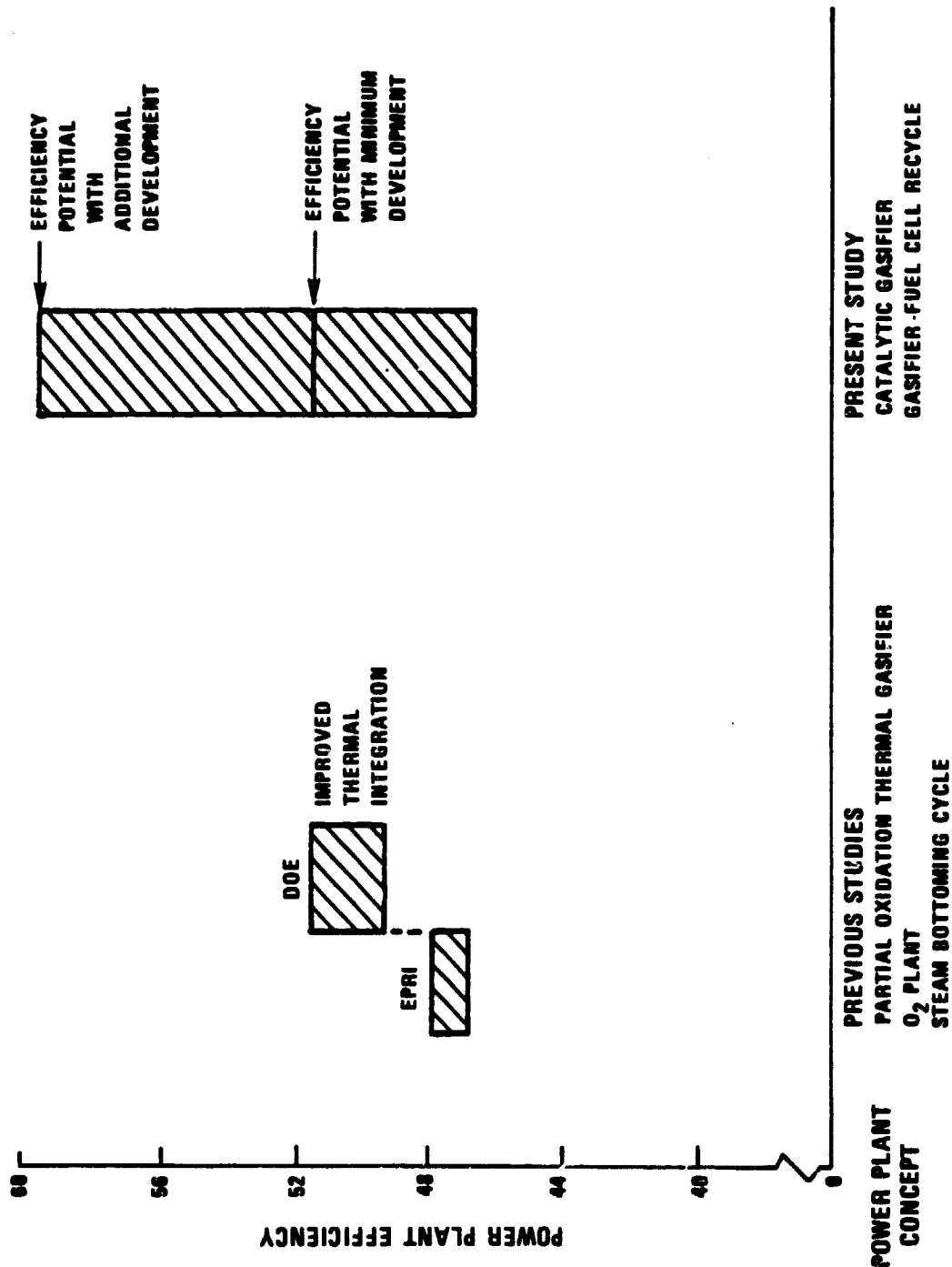


Figure 7-C. Efficiency Comparison for Range of Coal Gasifier-Fuel Cell Power Plant Concepts

TABLE 7-A

DEVELOPMENT PRIORITIES FOR CATALYTIC GASIFIER/
FUEL CELL POWER PLANT CONCEPT

FOR ALL POWER PLANT CONCEPTS

- o CONTINUED FUEL CELL TECHNOLOGY DEVELOPMENT WITH WASTE
HEAT REFORMING
 - o INTERNAL REFORMING
 - o SENSIBLE HEAT REFORMING

FOR SIMPLIFIED CONCEPTS AT 50-52% EFFICIENCY POTENTIAL

- o CATALYTIC GASIFIER DEVELOPMENT, SIMILAR TO EXXON GASIFIER
- o CHAR HEAT RECOVERY

FOR IMPROVED EFFICIENCY POTENTIAL

- o LOW TEMPERATURE CATALYTIC GASIFIER DEVELOPMENT
- o CHAR HEAT RECOVERY
- o HIGH TEMPERATURE DESULFURIZATION

1. Development of high temperature desulfurization, such as that being done at Morgantown Energy Technology Center by DOE.
2. Development of low temperature catalytic gasification.

Success in these two areas would greatly enhance the attractiveness of the integrated catalytic coal gasifier - molten carbonate fuel cell power plant with an efficiency potential approaching 60%.

Finally, in order to assure the correct design limits for carbon deposition, it is recommended to monitor gas handling experiences in future coal gasification and hydrocarbon processing.

8.0 REFERENCES

1. "Methane from Coal Aided by Use of Potassium Catalyst." Chemical Engineering, Vol. 89, No. 6 March 22, 1982, pp. 64-66.
2. "Catalytic Coal Gasification - An Emerging Technology." Science, Vol. 215, No. 4529, June 8, 1982, pp. 121-127.
3. Exxon Catalytic Coal Gasification Process Predevelopment Program, Contract No. E(49-18)-2369, Exxon Research and Engineering Company, Final Project Report to Department of Energy, December 1978.
4. Beasley, Brent E., et al. "Support of a Process Improvement on the Exxon Catalytic Coal Gasification Development Program," American Institute of Chemical Engineering 1982 Annual Meeting, Los Angeles, CA, November 14-19, 1982, Paper No. 119f.
5. "An Economic Comparison of Molten Carbonate Fuel Cells and Gas Turbines in Coal Gasification Based Power Plants," EPRI Research Project RP239-2, Final Report AP-1543, Fluor Engineers and Constructors, Inc., September 1980.
6. "Development of Molten Carbonate Fuel Cell Technology," Contract No. DE-AC01-79ET15440, Quarterly Technical Progress Report No. 7, April 1, 1981 - June 30, 1981, United Technologies Corporation.
7. "Assessment of Sulfur Removal Processes for Advanced Fuel Cell Systems," EPRI Research Project 1041-5, Final Report EM-1333, C. F. Braun and Company.
8. "Economics of Texaco Gasification - Combined Cycle Systems," EPRI Research Project 239, Final Report AF 753, April 1978, Fluor Engineers and Constructors, Incorporated.
9. "Development of Improved Molten Carbonate Fuel Cell Technology," EPRI Research Project 1085-4, Final Report, July 1983, United Technologies Corporation.